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The application of modern control systems to the operation of wastewater-treatment plants is discussed. Control strategies for the commonly used wet- and dry-weather treatment processes and their collection systems are described. Wherever possible, the benefits derived from, as well as the operating problems associated with, the actual or proposed control strategies are documented. Cost/benefit analysis indicates that many untried feed forward mass proportional control schemes are economically attractive because of the low payback periods. This study concludes that despite current concepts, the smaller (1 to 5 mgd) plants can afford and need significantly greater amounts of automatic control. A lack of reliable field-proven analytical sensors for important parameters appears to be the principal obstacle impeding the implementation of more sophisticated control strategies. Centralized control with semigraphic display should be used in treatment plants since it saves on operating labor, improves operation, and increases the safety of wastewater treatment. Automatic data acquisition systems are cost effective and should be used in medium and large sized plants. Direct digital control and computerized control can only be economically justified in large dry-weather treatment plants and large storm-water control networks.

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SELECTED APPLICATIONS
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IN
WASTEWATER-TREATMENT FACILITIES

by

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Contract No. 68-03-0144

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

Control strategies potentially applicable to wet and dry-weather wastewater-treatment facilities were evaluated during the course of this study. The evaluation included various levels of instrumentation and automation which could be utilized in the implementation of these control strategies. Cost/benefit analysis indicates that many untried control schemes are economically attractive because of the low payback periods. Furthermore, this study concludes that despite current concepts, smaller (1 to 5 mgd) plants can afford and need significantly greater amounts of automatic control. Finally, direct digital control and computerized control can be economically justified in large dry-weather treatment plants and storm-water control networks.

Francis T. Mayo, Director
Municipal Environmental
Research Laboratory

ABSTRACT

The application of modern control systems to the operation of wastewater-treatment plants is discussed in this report. Control strategies for the commonly used wet- and dry-weather treatment processes and their collection systems are described. Wherever possible, the benefits derived from, as well as the operating problems associated with, the actual or proposed control strategies are documented. Cost/benefit analysis indicates that many untried feedforward mass proportional control schemes are economically attractive because of the low payback periods. Furthermore, this study concludes that despite current concepts, the smaller (1 to 5 mgd) plants can afford and need significantly greater amounts of automatic control. However, a lack of reliable field-proven analytical sensors for all of the important parameters appears to be the principal obstacle impeding the implementation of more sophisticated control strategies.

Centralized control with semigraphic display should be used in virtually every treatment plant since it saves on operating labor, improves operation, and increases the safety of wastewater treatment. Automatic data acquisition systems are also cost effective and should be used in all medium and large sized plants. Direct digital control and computerized control can only be economically justified in large dry-weather treatment plants and large storm-water control networks.

This report was submitted in fulfillment of contract 68-03-0144 by Raytheon Company under the sponsorship of the U. S. Environmental Protection Agency. It covers a period from 30 January 1972 to 31 May 1975, and work was complete as of 31 May 1975.

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SECTION I

INTRODUCTION

During the past three decades, automatic process control has had a dramatic impact on most industrial processes. In particular, many chemical processes must be automatically controlled in order to turn out a satisfactory product; consider, for example, polyethylene production, where unsatisfactory control results in a reactor full of soot rather than high-grade plastic. Today's managers look toward instrumentation and automatic control as a means of increasing profits by:

- Maximizing productivity and minimizing off-spec products
- Reducing energy consumption and operating labor
- Automatically regulating the principal variables
- Computerizing supervisory control.

Despite the large number of successful applications of instrumentation and automatic control in the chemical, food, petroleum, and other related industries, the wastewater-treatment profession has been very slow to instrument and automate various wastewater-treatment processes. In fact, most treatment processes are designed as high-capacitance self-regulating processes that can be operated over a wide range of conditions without any loss in efficiency. This design philosophy evolved from the general lack of quick-time analytical instruments and automatic corrective control. For example, the activated sludge process in a municipal plant must accept wide influent flow rate and organic strength variations; accordingly, oversized aeration basins and secondary clarifiers that can absorb these variable loadings must be used. Although some wastewater-treatment organizations have attempted to catch up with the process industry in the use of instrumentation and control, much of the practical information (particularly as it relates to economic and technical feasibility) is not widely known.

One primary obstacle impeding a logical approach to instrumentation and control of wastewater-treatment plants stems from the lack of a comprehensive national state-of-the-art survey as a reliable source of baseline information. In one effort to plug this information gap, a team of experts in the areas of treatment process requirements, plant design, costs, analytical instruments, control devices, and computerized automatic techniques has performed a comprehensive study of the current state of the art. During the first phase of this project, the team conducted a survey of instrumentation practices in some 50 wet- and dry-weather wastewater-treatment facilities (1). In addition to this, all pertinent technical literature covering the base period of 1968 through 1973 was researched for instrumentation and automation information, and abstracts were provided for all of the pertinent literature (2).

The results of the user survey and literature review have been published in separate interim reports, thus, only the conclusions and recommendations will be repeated in this document. This report draws upon the survey's findings, published literature, case studies, and other pertinent information sources to develop the rationale for current and contemplated monitoring and control strategies for the commonly used wet- and dry-weather treatment processes. Instrumentation and automatic control devices contemplated for use in wastewater-treatment projects should, except for safety reasons, be justifiable on an economic basis. Accordingly, the use of a cost/benefit analysis as a decision-making aid is demonstrated. Much cost data on the application of instrumentation and automation for wastewater treatment and control are presented in this report for use by the reader in such cost/benefit analyses. The suitability of available instrument and automatic control devices is also appraised, and future research needs are reported.

The state of the art begins with a review of automatic measuring devices and analytical sensors that are needed to assess the pollution load and removal efficiencies. Then a basic philosophy of automatic open-loop, feedforward, and closed-loop control is presented. In the process control section, the principles of automatic control are brought to bear on the commonly used wastewater-treatment processes for both wet- and dry-weather facilities and their collection systems. For the reader's convenience, the process control section is organized on a unit processes and operation basis, with each subsection being virtually self-sufficient. Since the rapid advances of the control industry make centralized and computerized control very efficient, a separate chapter is devoted to this important topic.

After developing the process control strategies, the economic impact, payback periods, and annualized cost savings are presented in such a manner as to develop the criteria and demonstrate the most feasible procedures for finding cost-effective automatic control systems. In the final chapter, the principles developed are coupled with operating experiences to produce pragmatic instrumentation designs for 1 and 10 mgd biological and physical/chemical wastewater-treatment plants.

SECTION II

CONCLUSIONS

Based on this survey of 50 treatment facilities, nearly half the instrumentation employed in treatment facilities measures flowrates and liquid levels, whereas analytical instruments represent about only one-quarter of the current instrumentation used. Mechanical measuring devices dealing with position, speed, and weight account for the remainder of the instruments observed in treatment plants.

The following measuring instruments for on-line continuous service in wastewater-treatment activities are commercially available with sufficient reliability:

- Level
- Flowrate
- Temperature
- Pressure
- Speed
- Weight
- Position
- Conductivity
- Rainfall
- Turbidity
- pH
- Residual chlorine
- Free chlorine gas
- Hazardous (flammable) gas.

Sludge density meters, sludge blanket level detectors, on-line respirometers, dissolved oxygen (DO) probes, and many automatic sampling systems use well-established principles that are suitable for wastewater monitoring and control activities, but require significant maintenance.

The investigations revealed that most treatment facilities use considerably fewer instruments and automatic control devices than the closely related water supply and chemical processing plants. The cost and performance data collected during the survey of treatment facilities show that the average secondary plant allocates about

3% of its construction cost for installed instruments, whereas water supply and chemical treatment plants usually budget about 6% and 8%, respectively, for installed instruments. Remote satellite wet-weather treatment facilities, which in theory should operate unattended or at least with a minimal amount of operating manpower, budgeted only about 2% for automatic control, and they usually required manual assistance. Central computerized storm-water routing and in-line storage systems, however, employed an adequate number of instruments and automatic control devices, and performed satisfactorily.

The primary reason for the paucity of instrumentation use in existing wastewater-treatment plants is the unsatisfactory performance of many of the primary measuring elements and analytical sensors. Since most of the measuring elements interface directly with raw sewage, mixed liquor, or thickened sludge, these devices are subject to rapid fouling; accordingly, they need more frequent cleaning and calibration. Several on-line measuring devices for assessing the organic concentration (TOC, TOD, COD, and respirometry) of wastewater are commercially available at this time. These on-line analyzers require copious amounts of skilled maintenance, which is usually unavailable in most wastewater-treatment plants, and finally, many instruments have not been sufficiently evaluated in the field to form any conclusions.

Automatic control loops should be economically justifiable and cost effective. In this report the payback period presented as a function of plant size was used for the evaluation of automatic control loops. It is assumed that a payback period of 2.5 years or less is adequate to justify the purchase and installation of an automatic control loop although longer payback periods may also be cost effective. The cost/benefit analyses, which used actual field data wherever possible, show that most flowrate proportional, feedback, and combination feedback feedforward control strategies are economically attractive for plants in the 1 to 5 mgd range. In spite of the favorable economics of automatic process control, most wastewater-treatment plants use very little automation. The survey of 50 treatment facilities found that automatic chemical addition control, residual chlorine control, and digester temperature control were used by about only one-third of the visited plants. At these plants, most managers considered the automatic control systems cost effective since they do save energy and chemicals and, at the same time, improve the plant operation.

Central control of a plant organizes its operation so that all important events, alarms, and treatment information are displayed, and recorded in a centralized location. Virtually all of the large facilities surveyed successfully utilized central control. Moreover, a cost/benefit analysis shows that centralized control is economically justifiable in plants as small as 1 mgd. Most new plants use automatic data acquisition systems; approximately 20% of the visited facilities used data-logging computers. Compare this to the figure that only 10% of the plants used dissolved oxygen control. However, direct digital computer control is not well established in dry-weather treatment plants. Real-time computerized supervisory control of large storm and combined sewer networks is cost effective because the vast number of variables and control points exceeds human computational and decision-making capabilities.

Although the instruments and control systems in wastewater-treatment plants were adequately maintained, satellite storm-water treatment facilities were supplied with less than adequate maintenance. On the other hand, storm-water control centers that typically receive storm-water and combined sewer network information were well maintained and operated satisfactorily.

SECTION III

RECOMMENDATIONS

Cost/benefit analyses clearly show that most automatic control loops can be cost effective in smaller plants (i.e., 1 to 5 mgd). Therefore, greater amounts of automatic control should be specified for the following processes, even in small plants:

- Prechlorination
- Aeration
- Digestion
- Disinfection
- Phosphorous removal
- pH adjustment.

Intensive applications of elaborate and novel logic schemes, computers, displays, and recorders will not improve the effectiveness of wastewater treatment. Instead, well-documented field evaluation programs are needed to help appraise the potentially useful but untried control strategies developed in this work and elsewhere.

The need for more reliable operation to meet the quality standards and effluent permit requirements will encourage greater use of instrument and automatic control.

Nearly 70% of the plants included in a recent survey of 50 plants had neither operation nor cost data on their instrumentation. Wastewater-treatment plant management must make every effort to improve recordkeeping practices. Development of a uniform, easily practiced instrument recordkeeping system would facilitate improvements in this area.

Many misunderstandings and much confusion can be avoided in the future by having design engineers use standard symbols on instrument drawings.

Since instrument purchasing and installation are steadily becoming more complex, serious consideration should be given to new contractual procedures to ensure that specified instruments and control systems are operating and effective when installed.

Since instrumentation and automatic control devices require both maintenance and calibration, any plans for increased instrumentation must include plans for upgrading the qualifications of the maintenance staff.

Instrument technician training programs sponsored by the private sector and public funds should be expanded to combat the present-day shortage of qualified instrument technicians in the wastewater-treatment field.

Cost/benefit analyses that make use of total annualized costs should be used as decision-making aids when considering alternative and optional instrumentation systems.

Environmentally oriented professional societies and government research agencies, working together, should sponsor the writing of a comprehensive manual of recommended practices relating to the instrumentation and automation of wastewater treatment.

As a suggested guide for future research and development, the following list of sensors, control loops, and computer hardware and software sums up the most critical needs of wastewater-treatment instrumentation and automation:

- Sensors:
  - Rapid and automatic on-line organic monitoring devices
  - On-line wet chemical analyzers for ammonia and total phosphates
  - Flow meters for on-line use in sanitary, storm and combined storm systems.

- Control loops:
    - Organic load equalization
    - Food-to-microorganism ratio
    - Breakpoint chlorination
    - Mass proportional phosphate removal
- Computer hardware and software:
    - User-oriented language
    - Uniform data formatting and reporting
    - Standardized input/output requirements
    - Centralized software library containing program routines useful for wastewater treatment plant operation control and management.

SECTION IV

AUTOMATIC MEASUREMENTS

INTRODUCTION

Inasmuch as an automatic control system should respond to changes of the influent loading and ambient conditions, the automatic measuring devices that detect these changes are a significant part of the automatic control system. The analytical sensors and other measuring devices act as the eyes and ears of the control system and provide the input signal to the controller. The universal presence of uncertainty in any physical or analytical measurement must be clearly understood at the onset of an error analysis. These errors come from the device itself as well as from the standards used in calibrating the instruments. Discounting the significance of errors that arise from the calibration, the important sources of instrument errors in wastewater-treatment applications are:

- Noise that introduces a significant amount of spurious signals
- Response time or dynamic lag
- Interactions that cause a loss of measurement integrity
- Deterioration of the measuring device.

INSTRUMENT NOISE

A survey (1) of approximately 50 wastewater-treatment facilities has provided a large amount of data on the in-service history and operating experiences of many process-measuring instruments. Most of the output signals from dissolved oxygen (DO) probes, flowmeters, pH probes, and total organic carbon analyzers can contain an appreciable amount of noise in relationship to the true signal. Incomplete mixing, poor shielding, poor sampling, and poor instrument designs accounted for most of the noise observed in the on-line measuring devices. Fortunately, good engineering practices and simple filtering systems reject most of the unwanted noise components and, as a result, noise is not a significant problem with the current instrumentation systems used in wastewater-treatment projects.

RESPONSE TIME

Response time errors or dynamic lags may also contribute to the uncertainty of a measurement. If the sample characteristics are changing with respect to time, a significant lag or long response time results in a signal whose value depends on the time history of the sample over the previous interval of time. As an example, imagine that a cold thermometer, reading 20°C (68°F), is suddenly plunged into boiling water. The thermometer will not immediately read 100°C (212°F). In fact, a certain lag period is required to reach the 100°C mark. During the lag period the thermometer is producing an inaccurate measurement due to its dynamic lag. The instantaneous error depends on the actual temperature, how fast the temperature is increasing, and the characteristics of the thermometer itself. The difference between the actual and indicated temperatures becomes negligible only after steady-state conditions have been reached and maintained for a reasonable period of time.

Measurement lag is dangerous in automatic control systems because it can cause overshoots and process instability. Some analytical sensors that measure the composition of a flowing stream operate on a batch principle, which accentuates the response time problems. During the user survey, none of the observed automatic on-line analyzers—with the possible exception of those measuring respiration rates— had excessively long response times, taking into account the process time characteristics. Several of the analytical systems employed sampling networks which did introduce considerable lag times into the overall measurement system.

INTERACTIONS

Although interaction errors may arise from coupling and feedback to other measuring systems and power sources, interference by other chemical constituents is the most serious interaction problem in the use of analytical sensors in wastewater-treatment service. The reader should remember that analytical sensors are required to determine the trace concentration of a single component from a complex multi-component mixture. However, interaction problems should be largely diminished with the availability of better and more specific selective-ion probes.

MAINTENANCE

The physical or chemical deterioration of measuring devices, as well as other changes in their characteristics, may cause a corresponding change in their response time and indication; this also constitutes a significant error source. Since most measuring devices in wastewater service interface directly with raw sewage, mixed liquors, or thickened sludge, these devices are subject to rapid deterioration by fouling from solids deposition, slime buildup, and precipitation. Accordingly, they need frequent cleaning and calibration. Moreover, the operating experiences with measuring devices, which were accumulated during the user survey and are presented in Table 1, indicate that instruments used for wastewater applications require more maintenance than their industrial counterparts. The survey team found that most treatment plants supplied approximately 90% of the maintenance resources needed. Although most plants have reasonably well-qualified maintenance staffs, any plans that call for adding sophisticated instruments must also provide for upgrading the staff's qualifications.

COMMERCIALLY AVAILABLE, USEFUL INSTRUMENTS

Figure 1 indicates that flow and level devices account for nearly half the instrumentation employed in existing treatment facilities (1). Analytical instruments represent approximately a quarter of the instruments observed, while position, speed, weight, and other mechanical-type measurements total about 15%.

Table 1. INSTRUMENT PERFORMANCE

VARIABLE	INSTRUMENT	APPLICATION	TYPICAL COST	TYPICAL MAINTENANCE			RELIAB. (MTBF)	TYPICAL USE
				FRQ/YR.	MH/YR. STP	MH/YR. IND		
LEVEL	Bubbler	Tanks & Wet Wells	$200	12	8	4	1-2 yrs.	5-15 yrs.
	d/p Trans.	Digesters & Sludge	700	0.6	5	5	1-5 yrs.	5-15 yrs.
	Float & Cable	Tanks & Wet Wells	400	24	60	5	.2-2 yrs.	2-20 yrs.
	Optical	Sludge Blanket	1K	—	—	—	.1-5 yrs.	2-8 yrs.
FLOW	Flume & Weir	Major Flows	2K+	1.4	2	—	.5-5 yrs.	5-30 yrs.
	Venturi, etc.	Air and liquids	800+	4	20	6**	2 mo.-5 yrs.	5-30 yrs.
	Propellers	Clean liquids	1K+	7	10	10	1 mo.-1 yr.	1-8 yrs.
	Pos. Displace.	Gases	500+	2*	80*	10	1 mo.-1 yr.	1-5 yrs.
	Magnetic	Liq. and Sludge	2K+	12	12	8	.5-10 yrs.	5-20 yrs.
DENSITY	Nuclear	Med. & Thick Sludge	5K	48	51	40	1-3 yrs.	8 yrs.
	Mechanical	Med. & Thick Sludge	—	Excessive			1-6 mos.	2 yrs.
ANALYSIS	pH and ORP	Aqueous Liquids	2K	300	50	29	1-4 mos.	.5-5 yrs.
	Dissolved O_2	Aqueous Liquids	2K	100	60	—	1-9 mos.	.1-5 yrs.
	Res. Chlor.	Aqueous Liquids	5K	365	140	—	.2-1 yr.	4 yrs.
	Turbidity	Fairly Clean Liquid	3K	—	—	—	1-6 mos.	4 yrs.
	Conduct.	Aqueous Liquids	1K	200	60	—	1-4 mos.	4 yrs.
	Chlorine Gas	Airspace	3K	24	50*	—	.5-1 yr.	8 yrs.
	Explosive Gas	Airspace	3K	12	12+	50	.2-1 yr.	8 yrs.
	BOD, TOC, etc.	Wastewater	—	Excessive			.1-1 mo.	.3-1 yr.
MISC.	Temp.	All	300	1*	8*	1	.5-2 yrs.	5 yrs.
	Press.	All	200	5	4	4	.1-5 yrs.	5 yrs.
	Speed	Engines, etc.	—	—	—	—	.6-5 yrs.	5 yrs.
	Weight	Sludge or Cl_2	2K	24*	60*	—	.6-2 yrs.	10 yrs.
	Position	Sluice Gates	1K	18*	30*	—	.1-1 yr.	1 yr.
	Sampling	Liquid Streams	4K	0.5	20	—	.1-1 yr.	4 yrs.
	Rainfall	Storm Waters	500	24*	50*	—	1-5 yrs.	12 yrs.
CONTROL	Level	Wells & Basins						
	Flow	All Fluids						
	Sludge	Sludge Separation						
	Air Flow	Aeration						
	Dosage							
	Res. Chlorine	Chlorination						
	DO	Aeration						

NOTE:

STP = TREATMENT PLANT

IND = INDUSTRIAL, SEE TEXT

* Estimated

** d/p Converter only

Based on the field survey results (which are summarized in Figure 2), instrument users indicated that the following commercially available measuring instruments possess sufficient reliability for on-line use in wastewater-treatment facilities: level, flow rate, temperature, pressure, speed, weight, position, conductivity, rainfall, turbidity, pH, residual chlorine, free chlorine gas, and fire-hazardous (flammable) gas. Suspended solids analyzers may also be added to the list, based on the experience at Palo Alto, California (3).

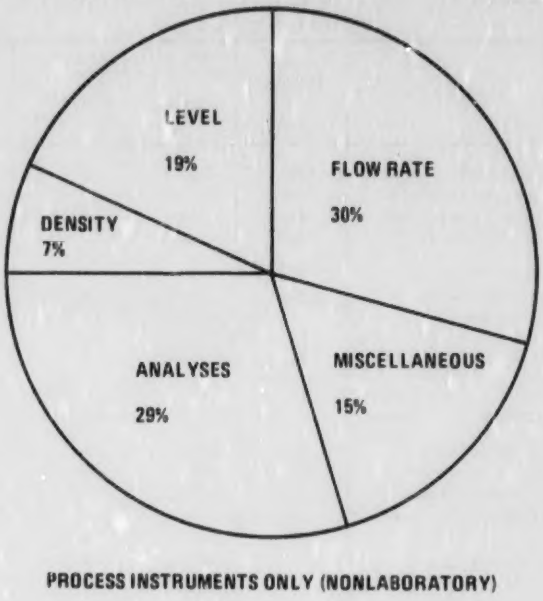

Figure 1. Distribution of measuring instruments observed during user survey.

Sludge density meters, sludge blanket level detectors, on-line respirometers, DO probes, and many automatic sampling systems use well-established principles that are suitable for wastewater monitoring and control activities but, based on the survey, require significant maintenance. These instruments need improved maintenance characteristics before they will become more widely used.

INSTRUMENT NEEDS

In spite of the many successful flow-measuring devices used in treatment plants, the accurate and reliable monitoring of storm-water flow poses special problems. High transient flows, large operating ranges, high suspended solids, and frequent collisions with large debris are merely some of the obstacles that an acceptable in-sewer storm-water flowmeter must overcome. Consequently a suitable storm-water flowmeter needs to be developed that will accurately produce the flow rate data required for sewer regulation.

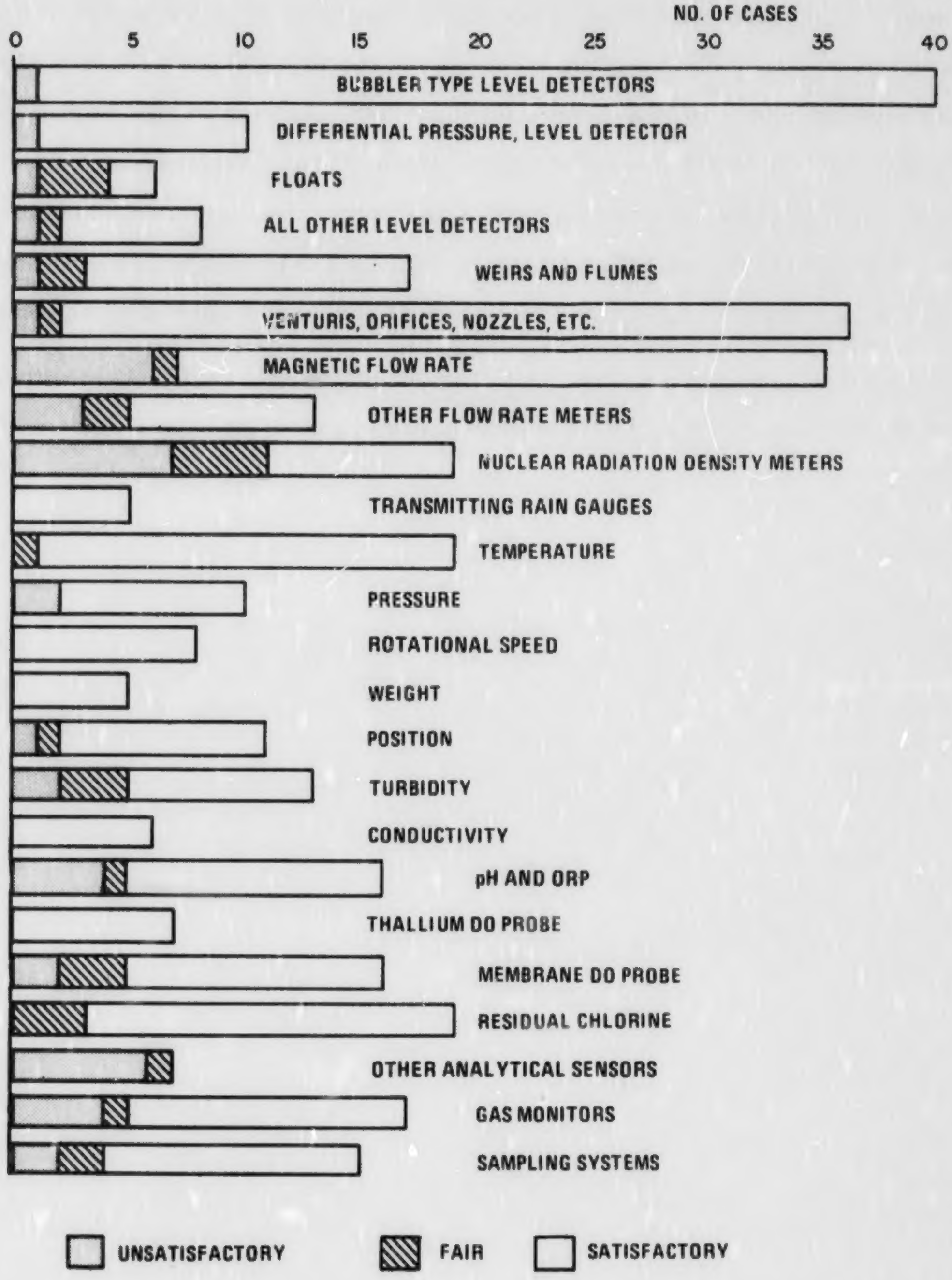

Figure 2. Performance summary for measuring devices in wastewater treatment.

Some of the most important process parameters (such as total organic carbon and suspended solids) have never been successfully monitored on an automatic basis in wastewater-treatment plants. Current studies at the EPA-DC Pilot Plant at Blue Plains have indicated that Dohrmann-Envirotech TOC analyzer and the Biospherics suspended solids analyzer are satisfactory. Work at Palo Alto (3) showed that the Biospherics and the Keene suspended solids analyzers were comparable in performance. If treatment process efficiency and reliability are to improve, then the following analytical sensors—as yet unavailable—are needed to provide real-time control:

- Rapid and automatic on-line total organic analyzers or substrate monitors
- On-line wet chemical analyzers for ammonia and all phosphorus forms.

SECTION V

AUTOMATIC CONTROL

Since the flow rate and strength of raw sewage fluctuate diurnally and seasonally, wastewater-treatment plants operate under variable loading. This may be the prime cause of effluent quality variation. Storm events and oil, grease, or industrial chemical dumps also upset wastewater-treatment plants. To minimize the impact of these deleterious disturbances, most treatment plants utilize high-capacity self-regulating processes that may unnecessarily increase capital and operating costs. However, major failures such as biomass washout and digester souring still occur all too frequently. Most investigators believe that the thoughtful application of automatic process control will improve the operation, maintenance, efficiency, productivity, and reliability of most wastewater-treatment facilities.

Ideally a control system should accomplish its desired objectives without incurring any errors; in fact, this is impossible because our process knowledge and control actions are imperfect. All that one can expect to accomplish is the desired task, with a minimum amount of disturbances. The goals of wastewater-treatment plant automation are as follows:

- Improve treatment reliability
- Reduce operating and maintenance costs
- Minimize effluent variability
- Detect problems and institute corrective measures
- Reduce equipment and structure sizes through increased productivity

- Find control actions that are readily adaptable by new and existing treatment plants
- Devise stable control systems.

Undoubtedly very few existing control schemes could simultaneously satisfy all of the above goals but, with good engineering judgement, the best mix of controller goals should guide the design engineer.

Inasmuch as most personnel associated with wastewater treatment do not have background in automatic control theory, a brief explanation of automatic control, its various modes, and important terminology follows. An automatic control system consists of the following:

- The process
- Instruments, sensors, flow-measuring devices, and an automatic analytical device
- Transmission devices to carry the instrument signals
- Controller logic to implement the strategy
- Final control elements to execute the control strategy.

Most investigators view process control as an automatic adjusting of the manipulation variable so as to maintain the desired balance in the process output stream. The controller modes can be classified (in the order of their complexity) as:

- Open-Loop Control—This simply involves estimating the form or number of actions necessary to accomplish a desired objective. It is a predictive controller, since no check is made to determine whether or not the corrective action taken has accomplished the desired objective. Figure 3 shows a typical open-loop control system. Open-loop control is capable of perfect control; however, if any one of the variables affecting the desired outcome deviates from its predicted values in either quality or quantity, open-loop control will not give perfect control. Accordingly, open-loop control is suitable only for well-known highly repeatable processes such as mechanical equipment or time-phase events.

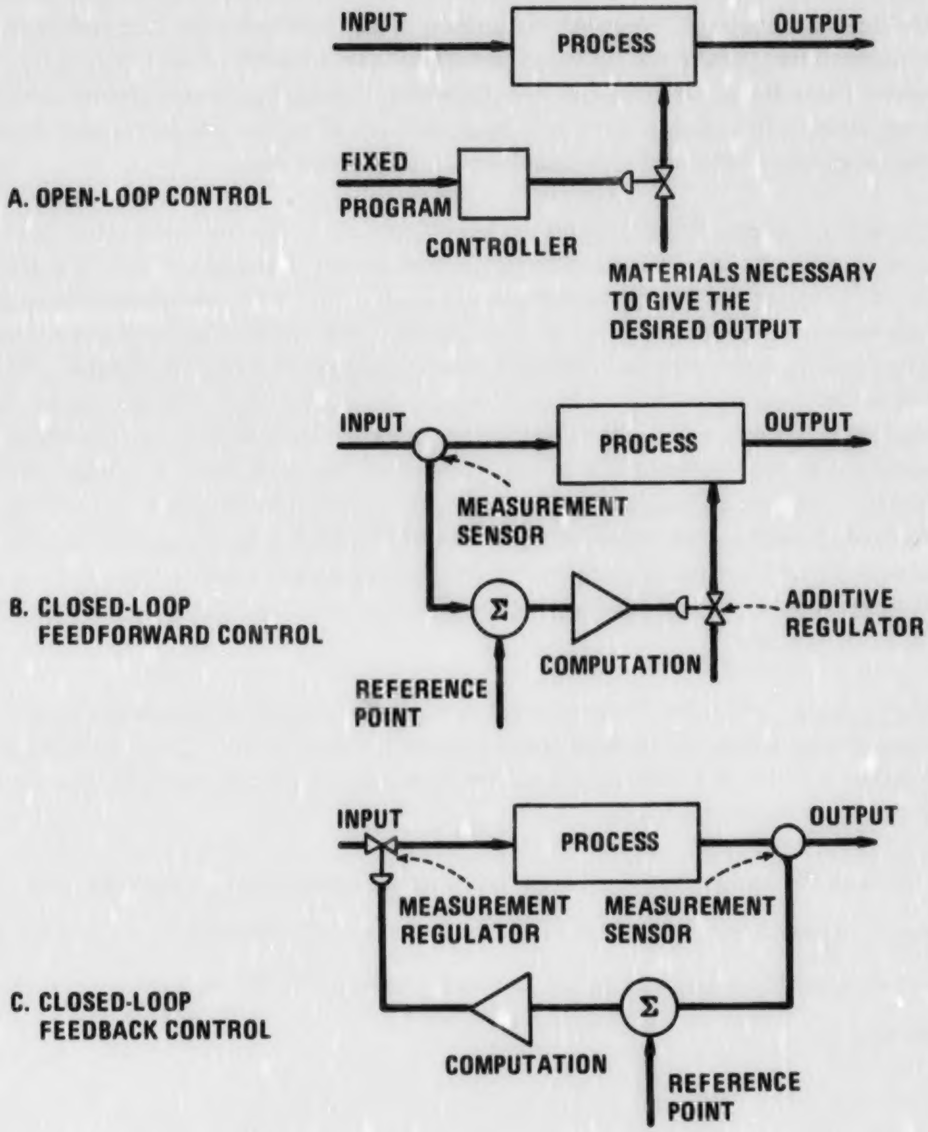
INPUT
PROCESS
OUTPUT
FIXED
PROGRAM
CONTROLLER
MATERIALS NECESSARY TO GIVE THE DESIRED OUTPUT
A. OPEN-LOOP CONTROL
INPUT
PROCESS
OUTPUT
MEASUREMENT SENSOR
Σ
ADDITIVE REGULATOR
B. CLOSED-LOOP FEEDFORWARD CONTROL
COMPUTATION
REFERENCE POINT
INPUT
PROCESS
OUTPUT
MEASUREMENT REGULATOR
MEASUREMENT SENSOR
C. CLOSED-LOOP FEEDBACK CONTROL
Σ
COMPUTATION
REFERENCE POINT

Figure 3. Basic types of process control loops.

- Closed-Loop Feedforward Control— One of the newest and most potentially useful control modes, this has often been referred to incorrectly as open-loop control. As shown in Figure 3, feedforward control is based on input measurements and computations to predict the amount of additive necessary to produce the desired product, whereas open-loop control relies only on predetermined assumptions. Since the process variables are monitored and corrective actions are taken continually, feedforward control is really a form of closed-loop control in which a complete loop is formed by the measurement device, the controller, the additive regulator, and the process.

- Closed-Loop Feedback Control— In feedback control, the controlled variable is measured and then compared to the reference or desired value. If a difference or error exists between the actual and desired levels, the automatic controller will make the necessary corrective adjustments in the final control element. Figure 3, which contains a typical feedback control loop, illustrates the differences between open-loop, closed-loop feedforward, and closed-lopped feedback modes. Feedback control depends only on the repeatability of the controlled variable measurements and the suitability of the controller's design and adjustments. No prior or predictive knowledge or assumptions are needed for feed control systems. Because of the technical and economic impracticability of precisely predicting the number of corrective actions needed to achieve satisfactory results with open-loop or feedforward control, feedback control is used most frequently.

- Multi-Loop Control— When several open-loop, closed-loop feedforward, and closed-loop feedback control loops are united in a single control strategy appropriate for the pertinent process requirements, the result is known as a multi-loop control system.

Sections VI and VII make use of all four types of automatic control systems in developing process control strategy; and, in Section VIII, the merits of the alternate control strategies are reviewed in the context of achieving the most cost-effective control scheme.

SECTION VI

CONTROL OF DRY-WEATHER TREATMENT PROCESSES

INFLUENT PUMPING AND PRETREATMENT CONTROL

Raw Sewage Pumping

The raw sewage system consists of a pumping station, with either wet and dry wells or submergible pumps in a wet well. As the sewage flows into the pumping station and the level rises, the pumps begin in sequence so as to pump the station down to a nearly empty condition. The control system described herein is for the start/stop control of two pumps with automatic alternation.

Although the measurement of flow into a pumping station is usually quite difficult due to the piping configuration, the measurement of flow out of a pumping station is relatively simply (by using any of the common flow elements). In this case the instrumentation consists of a magnetic flow element, a transmitter, a telemetry system, a flow recorder, and a flow totalizer.

Since pump control normally is based on the level, a level transmitter utilizing a bubble tube or diaphragm transmits its signal to a level controller containing a contacting mechanism that adjusts the start/stop point on both pumps. It is common to have both pumps start at different levels on a rising level in the wet well and then have them continue to operate until they reach a common shutoff point.

Normally the pumps are sized so that a single pump can accommodate all but the most extreme flows anticipated; even large flows may be handled but, in consequence, the pump downtime is lengthened. As a result, the second pump is essentially a backup pump in case the first pump fails or it becomes necessary to remove one of the pumps for service and/or repair. If these pump control systems are wired permanently into specific contact points, the first pump will be used much more than the second one and usually will wear out within a few years. This problem is simply resolved by using an automatic alternator, with each sequence starting on a rising level and terminating with a common shutdown on a falling level. A simple ratchet relay can be used to alternate the lead pump's position at the end of each cycle. As shown in Figure 4, hands-off automatic switches are located on each pump so that either one can be removed from service or operated manually if required.

There is an almost infinite number of variations possible in this type of station, the most common being to use a proportional-type level controller in conjunction with a variable-speed motor-driven pump. Essentially this permits the flow going out of the pumping station to match the flow coming in. The practical application of these level control techniques, however, depends on the hydraulics of each station.

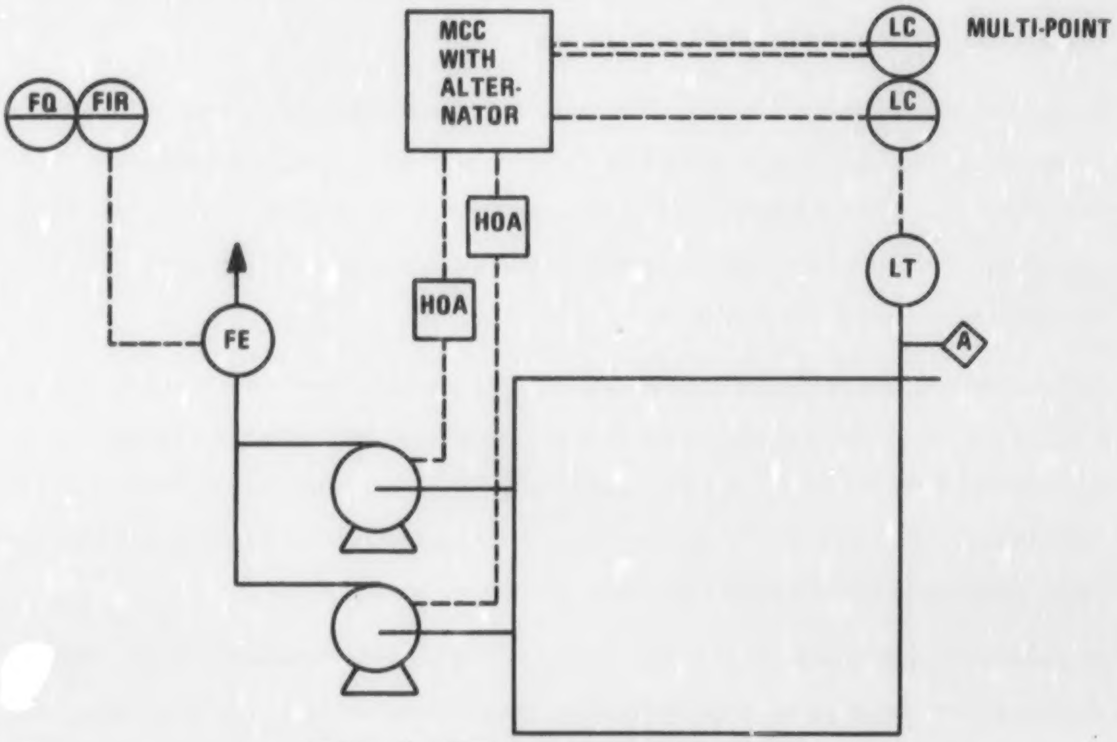

Figure 4. Raw sewage pumping control.

Screening

The purpose of screening is to remove large floating objects that might otherwise damage the plant's machinery. Inevitably a great many organic solids are picked up in the screening process.

As the screen becomes loaded with accumulated materials, a differential liquid level develops between the upstream and downstream faces of the screen. This differential level can be used to actuate control of the screen's cleaning mechanism. Using the differential level makes the system unresponsive to the absolute level within the channel, which can vary widely and independently from the differential level. It is important to note that the cleaning cycle should be so interlocked with associated devices that, once the cycle has been initiated, it must be completed in one operation. The level measurement is accomplished by means of bubble tubes located on opposite sides of the screen. Figure 5 illustrates the associated instruments and control devices.

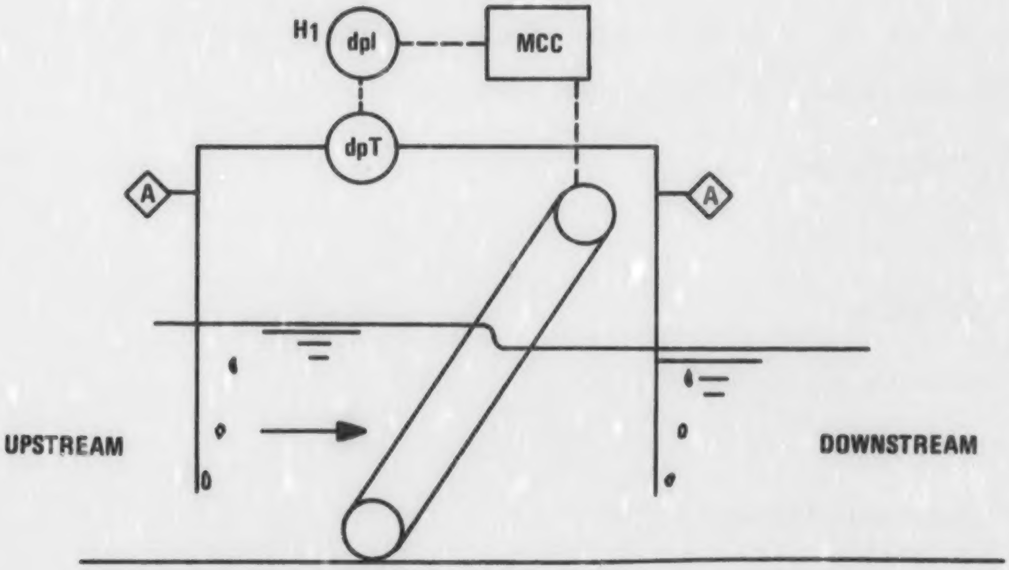

Figure 5. Screen control.

Variations of this arrangement have been in use for years. The same fundamental principle was followed in rather crude mechanical devices, consisting of heavy cast-iron floats and chain arrangements in float wells located on either side of the screen.

PRIMARY SLUDGE PUMPING

Introduction

The treatment of wastewaters by gravity separation is among the oldest methods used for solids/liquids separations. Decades of primary treatment plant operation have produced much practical knowledge and many "cookbook" operating techniques that lead to effective removal of the settleable solids in sewage. Approaches to successful treatment plant operation must consider the following:

- The specific gravity of raw sludge is only slightly greater than 1.0.
- Efficient sludge collection for subsequent treatment and disposal requires that the minimum possible quantity of water be removed with the sludge.
- Sludge must be removed continuously or frequently in order to minimize septicity and excessive loading on removal mechanisms.

The process dynamics of primary sedimentation are relatively simple as far as control is concerned. The basic representation is as follows:

$$S_L = K(Q_I, S_s, \%_R) \tag{1}$$

where:

Q_I = influent rate

S_s = settleable solids concentration

$\%_R$ = % removed

S_L = sludge settled formation rate

K = units constant

Equation (1), which represents the sludge formation rate, shows that over long periods the sludge formation rate will equal the removal rate. However, these rates are not balanced over short intervals, since sludge is usually settled and allowed to compact in the clarifier before being removed by pumping. For a reasonable period of time (e.g., 12 to 24 hours), the following applies:

$$S_p = S_L \qquad (2)$$

where:

S_p = sludge pumped

The principal operating objective of primary clarifiers involves keeping the removed sludge above some minimum density level, which is usually expressed as percent solids. This ensures the maximum practical capacity for subsequent treatment, since excessive quantities of water overload sludge treatment facilities and seriously impair their efficiency. This relationship is represented by the following:

$$D_s \geqslant \text{minimum value} \qquad (3)$$

where:

D_s = sludge density (% solids)

The relationships and practices discussed in this section have been either a common practice or a common operating objective for so long that current technical literature, with the exception of teaching texts, no longer deals with this process phase. However, the implementation of these common operating practices by means of automatic instrumentation and control equipment is recent. The key to successful control is the availability of reliable measuring and control instrumentation.

Since the only final control elements available are the raw sludge pumps and diversion valves, the control modes for raw sludge pump are limited to on-off and variable-speed sludge removal. Most of the small-and medium-size installations that use pumps rather than diversion valves operate these pumps on an on-off basis, since variable-speed pumping can result in low line velocity which favors clogging. Consequently, the presently available control techniques are not especially enhanced by the availability of variable-speed pumps as opposed to the on-off operation of constant-speed pumps.

Percent of Time Control

The simplest and most commonly used method of sludge withdrawal control is the operation of the sludge pumps, based on a percent of time. This can be expressed as follows:

$$Sp = f(t) \tag{4}$$

where:

t = time

In practice a percentage timer or interrupter is normally used. Such a device usually has a fixed cycle length (e.g., 60 minutes), and the percent of on-time is adjustable. On the basis of operational tests and practical experience, the plant operator sets the percent of on-time (e.g., 20%). The 60-minute cycle would result in the raw sludge pump's operating 12 minutes out of each 60 minutes. The pump would run for 12 minutes, shut down for 48 minutes, and then repeat the cycle.

Since the percent-of-time control scheme is an open-loop control strategy, variable influents, settling conditions, and the behavior of the other sludge removal equipment (as well as other factors that impact sludge density) will change the desired duty cycle. The percent of on-time would then be manually adjusted up or down to produce the desired concentration of solids in the raw sludge. The simple percent-of-time control strategy suffers from being insensitive to plant equipment and process conditions, and requires frequent operator attention in order to produce a high-quality sludge.

Percent of Time Control With Flow Proportion

If the settleable solids content of the influent sewage were constant, then withdrawing settled sludge at a rate proportional to the influent flow would be adequate, since this method overcomes the influence of the largest variable flow. However, it does not consider variations in settleable solids concentrations of the plant influent, changes in the other sludge-processing equipment, and process upsets in the primary clarifier.

Because the implementation of flow-proportioned sludge withdrawal normally involves operating the sludge pumps at variable speeds, incipient line settling and plugging can be overcome by using the pulse duration pump control. With this technique the flowmeter is equipped with a cam switch that has a fixed cycle (e.g., 30 seconds). The portion of the 30-second cycle when a circuit is closed is proportional to the flow measurement; for example, at 50% of flow span, the circuit will be closed for 15 seconds and open for 15 seconds. Since these time periods are too short for pump operation, the amount of on-time must be accumulated on a timer which usually is set for a time representing a fixed volume (e.g., 100,000 gallons). When sufficient time (wastewater volume) has accumulated, a second timer is actuated to operate the sludge pumps for a fixed period.

If the variation in waste flow is wide, it may be desirable to combine time and flow proportional control. This would be accomplished by having the sludge pumps operate on flow proportional control during the day and early evening. During the extremely low flow period at night, the sludge pumps would operate on straight timer control, since a straight flow-proportioned control would allow the clarifier sludge to become septic.

To our knowledge, the flow proportional method of sludge pumping is not in service in any existing facility. Nevertheless, the method has merit for small installations where more elaborate devices requiring continuing maintenance and skilled operations would probably fall into disuse.

Feedback Sludge Density Control

The commercial availability of radioactive sludge density meters opened new horizons for sludge pumping control. Although the application of radioactive source density measuring devices has become quite common, service and maintenance problems can be difficult. With proper installation and a reasonable preventive maintenance program, however, radioactive sludge density measuring devices provide useful information and may be included in an automatic control loop.

Feedback sludge withdrawal control uses the actual sludge density to regulate pumping so as to ensure that the density is above some minimum value. Sludge pumping control can be accomplished by using a combination of timers in conjunction with the density measuring devices, as shown in Figure 6. The sequence of operation is as follows:

- The raw sludge pump is started at a preset interval determined by a cycle timer.
- A second timer holds in the pump control circuit until a fresh sample of sludge is in the density element pipe section.
- The raw sludge pump will continue to operate until it measures a density that is below a preset minimum value. At this point, the sludge pump is stopped.
- The cycle is repeated at regular intervals, as determined by the cycle timer.

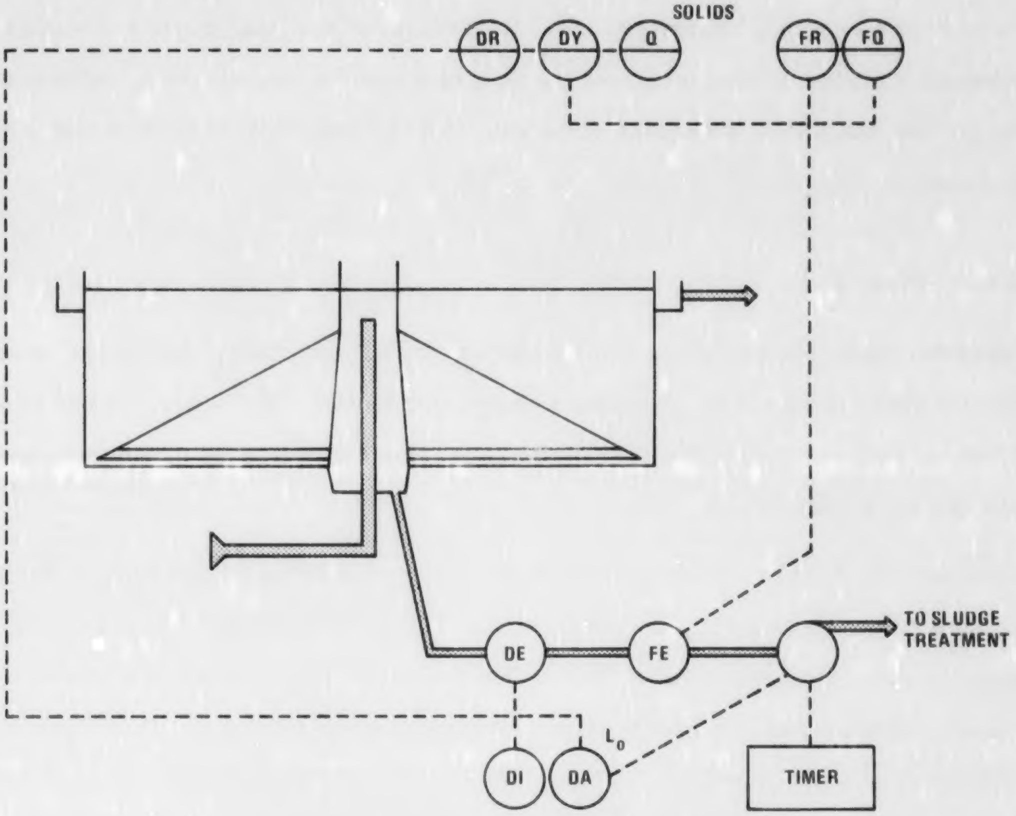

Figure 6. Sludge density control.

Feedback sludge density control offers the following advantages:

- Withdrawn sludge does not fall below the predetermined minimum density.
- Some sludge is pumped at regular intervals, which helps to minimize the development of septic sludge in the clarifier.
- Excessive sludge loading cannot build up to the point where the sludge removal mechanism would be damaged.
- Sludge pumping control automatically reacts to changes in the influent, the clarifiers process conditions, and the sludge removal equipment.

The importance of minimizing the waste sludge cannot be overemphasized. Since the volume of sludge varies as the percent solids and not percent liquid fraction, a 1% change in percent solids will have a dramatic impact on sludge volume. For example, a change in solids at raw sludge from 3% to 4% (97% to 96% water) will result in a volume reduction of approximately 33%. Regardless of the final method of sludge treatment, the size of equipment and the cost of operation are adversely influenced by a sludge that has a low percentage of solids. This is especially true in the case of digesters.

Interface Control

To prevent sludge accumulation and subsequent solids carryover, the sludge wastewater interface can be used to regulate sludge withdrawal. In the past, operators did this manually by noting the staining of a piece of Turkish towel which they dipped in the clarifier with a pole.

The recent development of an optical technique for measuring the position of the sludge waste interface has made this a viable automatic control technique. A sensing device consisting of a light source and a sensor that are a few inches apart is mounted on a carriage, which is motor driven vertically so that it passes through the possible area of sludge wastewater interface. Passing through the interface breaks the light path, and the position of the sensor at that moment corresponds to the top of the interface or the position of the settled sludge.

The availability of periodic measurements of the settled sludge interface permits automatic feedback control of sludge pumping, as illustrated in Figure 7. The sludge pumps start on a fixed cycle and shut down on the basis of the sludge water interface falling below a preset limit. This technique is essentially the same as that used with density measuring devices, but it ensures that the sludge blanket stays within a safe region of the primary clarifier.

To effectively regulate withdrawn sludge density as well as the sludge blanket depth, the sludge density and sludge interface control schemes can be combined. If a conflict exists (such as a rising blanket and too low a sludge density), then an interlock control system can alert the operator so that he can decide the course of action to take.

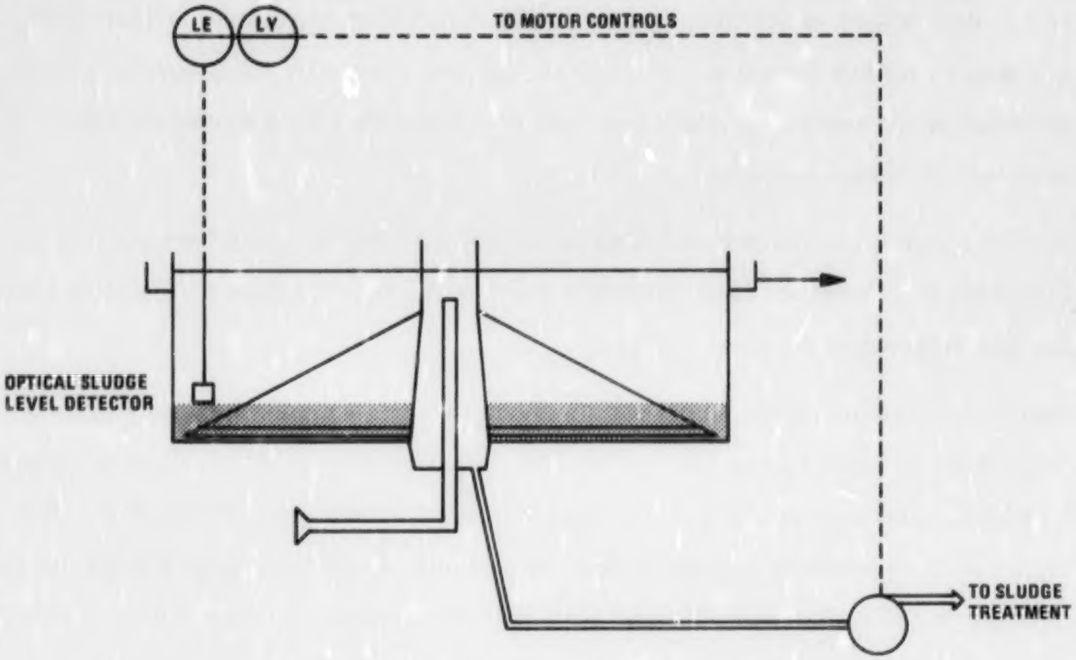

Figure 7. Sludge interface control.

Application Notes

These control systems are all highly stable because they all operate as successive batching techniques. Inasmuch as any system or components thereof can fail, the impact of such failures must be examined. Automatic sludge pumping failures fall into the following three general groups:

- Failures resulting in the complete suspension of sludge pump operations.
- Failures resulting in the partial suspension of sludge pump operations; such failures occur intermittently and are due principally to primary element failures.
- Failures resulting in continuously operating sludge pumps; these failures, too, are due principally to primary element failures.

A complete failure of the first type is usually detected by the plant operator before any major problem develops. If problems do occur, however, they consist usually of overloading the scraper mechanisms, which can be detected by the motor load detectors or torque switches.

Partial failures are much more difficult for the operator to detect because they are intermittent. However, they ultimately produce the same results as complete failures and are detected in the same manner.

Shutoff failures are the most important, since they can completely upset the sludge treatment and disposal system. In the case of sludge digesters, the digester becomes overloaded and digester failure can soon follow. The only practical method of detecting a failure wherein the sludge pumps run continuously is to arrange a timer so that an alarm sounds whenever a normal period of pump operation is exceeded. Fortunately, shutoff failures are rare.

Automatic control of raw sludge pumping permits the process regulation of primary clarifiers, as summarized in Table 2. These methods provide an economical solution to the critical problem of properly controlling sludge pumping and represent a marked improvement over manual operation, since they minimize the costs of pumping, dewatering, and digesting sludge.

CONTROL OF DISSOLVED OXYGEN IN AERATION BASINS

Today most secondary treatment facilities use aerobic biological processes, such as conventional activated sludge, step aeration, completely mixed, and contact stabilization, to remove soluble and colloidal organic pollutants. One of the most important requirements for achieving high BOD removals is maintenance of proper dissolved oxygen (DO) levels in the aeration vessels. If the DO drops below a critical level the effluent quality deteriorates. Excessive DO concentrations can hinder secondary

Table 2. SUMMARY OF PRIMARY SLUDGE PUMPING CONTROL STRATEGIES

Control Method	Benefits and Potential Savings	Advantages	Disadvantages
Time cycle	Better separation and 10 to 30% labor savings over manual control	Simple, inexpensive, reliable	Depends on experience and judgement for adjustments (can pass sludge in overflow and water in underflow)
Time cycle with flow proportion	Same as above	Reliable and responsive to flow rate changes	Requires flow signal and more instrumentation than time cycle alone
Sludge density with auxiliary timers	20 to 50% less labor than manual operation; reduction of water pumped with sludge can extend sludge handling capacity	Good, reliable separations; densities measured and consistently controlled; most free water kept out of sludge	Unrealistic repair delays and downtimes, although basic instrument is good; expensive; some water drawn off with/sludge
Sludge level with auxiliary timers	Same as above	Good, reliable separations; sludge level maintainable at best level; low cost	Procedure not fully developed; probe failure rate high and should be better; influence partly by color and turbidity

solids settling and consume large amounts of energy (4). Experience shows that a DO level of about 1.0 mg/l promotes process stability, high BOD removals, and minimum aeration energy consumption.

Since the microorganisms consume oxygen according to their metabolic requirements, the mixed liquor must be aerated to maintain a satisfactory DO level. Diffused aerator devices and mechanical aerators are frequently used to transfer oxygen to the mixed liquor, and power consumption is directly proportional to oxygen demand.

To maintain a constant DO level the oxygen supply rates should match the oxygen uptake rates. Because the uptake rates change continually due to variations in the organic loads, microorganism concentration, and degree of nitrification, it is difficult and time consuming to manually adjust the aeration equipment. Instead, most plants transfer more oxygen than necessary to save manpower and to ensure an adequate DO level.

Automatic DO control systems can modulate the rate of oxygen supply to the mixed liquor by adjusting the aeration equipment's power consumption. A good DO controller should hold the DO at the prescribed level in spite of changing loads and ambient conditions. The control system also needs to be stable and easy to maintain. A nationwide survey of instrumentation experiences (1) found that 20% of the secondary treatment plants used automatic DO control. Of the users about 70% were satisfied with automatic DO controller performance, and they reported power savings from 10 to 40% and BOD removal increases of about 10%.

The subsequent text describes automatic DO control systems that have been successfully practiced in wastewater-treatment plants and some promising—but untried—control strategies. Application notes and instrument diagrams are also included.

DO Control Systems

Aeration basins are classified as plug flow or completely mixed, based on flow patterns, geometry and mixing characteristics. Plug flow vessels (large length-to-width ratios) are used in the conventional activated sludge, step aeration, and contact stabilization processes, whereas the completely mixed activated sludge processes modification is practiced in rectangular or circular vessels. Aerated lagoons, which are popular for industrial wastewater treatment, exhibit mixing characteristics that fall in between plug flow and complete mix. Because of the difference in contacting patterns and oxygen consumption rates, DO control strategies depend on the type of aeration basin and the activated sludge process.

Inasmuch as most new treatment plants use the completely mixed activated sludge process and plug flow basins behave like several completely mixed vessels in series, the following text explains the rationale behind DO control schemes for a completely mixed aeration basin. Moreover, a diffused air system is used in the strategy development. Extensions to plug flow aeration basins and aerated lagoons, as well as to other types of aeration equipment, follow directly.

A material balance around a completely mixed aeration basin gives a relationship between dissolved oxygen concentration, influent flow rates, respiration rate, and airflow rates:

$$V\left(\frac{dC_o}{dt}\right) = FC_i - FC_o + kl_a\,(C_s - C_o)\,U\,V - R\,V \qquad (5)$$

where:

C_o = aeration basin DO concentration (mg/l)

C_i = influent DO concentration (mg/l)

F = flow rate (liters/min)

kl_a = transfer coefficient $(\text{liters})^{-1}$

U = airflow rate (liters/min) at standard conditions

C_s = saturation value of DO at system temperature and pressure (mg/l)

R = respiration rate or oxygen utilization rate $(\text{mg/l min})^{-1}$

V = volume of aeration basin (liters)

Rearranging terms, the oxygen balance becomes:

$$\frac{dC_o}{dt} = kl_a\,(C_s - C_o)\,U - R\,\theta + \frac{C_i - C_o}{\theta} \qquad (6)$$

where:

$$\theta = \frac{V}{F}$$

The respiration rate, R, is related to the substrate utilization rate and the endogenous respiration rate. Eckenfelder (5) expressed this relationship by the following equation:

$$R = \alpha \dot{S}_r + \beta X_{vss} \tag{7}$$

where:

α = coefficient relating substrate removal to oxygen consumption

$\dot{S}_r$ = substrate removal rate

β = endogenous constant rate

X_{vss} = mixed liquor volatile suspended solids

A substrate balance around the aeration basin yields a relationship between loadings and removal rates:

$$\frac{V\,dS_o}{dt} = FS_i - FS_o - \dot{S}_r V \tag{8}$$

Several kinetic models such as two phase, second order and Monad (Michaelis-Menton) have been proposed to describe substrate removal rates. In general, the reaction rate increases with substrate concentration. This control system analysis, however, will not depend on a particular kinetic model.

Equations (6), (7) and (8) were consolidated into the following overall relationship in terms of the fundamental parameters:

$$\frac{dC_o}{dt} = kl_a\,(C_s - C_o)\,U - \alpha \left[\frac{(S_i - S_o)}{\theta} - \frac{dS_o}{dt}\right] + \theta\beta\, X_{vss} + \frac{C_i - C_o}{\theta} \tag{9}$$

Rearranging Equations (6) and (9) in such a manner as to give the airflow requirements as a function of system variables yields Equation (10):

$$U = \begin{cases} \left[R\,\theta + \dfrac{C_o - C_i}{\theta} + \dfrac{d\,C_o}{dt} \right] \dfrac{1}{kl_a\,(C_s - C_o)} & \text{(10a)} \\[2ex] \left[\dfrac{\alpha}{\theta} \left(S_i - S_o - \dfrac{d\,S_o\,\theta}{dt} \right) + \beta X_{vss}\,\theta + \dfrac{C_o - C_i}{\theta} + \dfrac{d\,C_o}{dt} \right] \dfrac{1}{kl_a\,(C_s - C_o)} & \text{(10b)} \end{cases}$$

The airflow rate requirements, according to Equation (10a), increase with respiration rate, dissolved oxygen concentration, and the rate of DO change in the basin. Since ambient and wastewater characteristics change C_s and kl_a, air requirements also vary according to these factors. Equation (10b) shows that increased substrate loading and higher mixed liquor volatile suspended solids also raise the airflow requirements. Similar equations and conclusions are obtained for all types of aeration devices and process configurations but, instead of air requirements, the energy consumption changes as a function of the previously mentioned parameters. These findings agree with operational experiences observed in activated sludge plants and aeration lagoons.

Many manufacturers supply oxygen transfer equipment; this equipment is broadly classified as diffused air devices, mechanical aerators, or combinations of both. Diffused aeration systems are further categorized, based on the type of dispersion media as porous diffusers (fused aluminum oxide) or non-porous diffusers (spargers, ridge and furrow). Updraft, downdraft, plate and brush are the major classes of mechanical aerators. For all types of aeration equipment, the power consumption is directly proportional to oxygenation output.

Selecting the manipulation variable involves examining the input/output relationships of the aeration equipment for the purpose of attaining the control objectives. Available final control elements which execute the control strategy also influence the choice of the manipulated variable. Specifically, the manipulation variable must not only regulate the DO level, but it must reduce energy consumption as well. For example, consider a centrifugal blower with operating characteristics, as shown in Figure 8. Further assume this blower is being driven by a single-speed induction motor. Clearly, airflow rate under standard conditions (equivalent to mass rate of air, kg/min) is a prime manipulation variable; however, the point of application and the implementation method are crucial. Throttling the discharge line will reduce the DO in the aeration basin, but little or no power saving will be obtained since the throttled valve dissipates the excess energy. Consequently, throttling the discharge of a centrifugal blower is an unacceptable point of application. To achieve good DO regulation and to effect power saving, the inlet (low pressure) line should be throttled. If the blower was equipped with inlet guide vanes, adjustment of these vanes would result in even greater power savings. The final control element would be a valve with an automatic operator for the throttling case and an automatic positioner for the inlet guide vanes case. Both methods would regulate airflow to the inlet side of the blower.

As a second example, consider lobe-type blowers. Since these devices are designed to work over a relatively narrow pressure range, throttling the inlet or discharge will not effectively reduce the airflow rates, and the power consumption will remain about the same. Lobe blowers, consequently, should be equipped with multiple-speed driving motors for good DO control and power savings. However, in large plants that use four or more blowers, simply turning the blowers on or off may be a suitable method for regulating airflow rates and single-speed motors would suffice. A certain minimum power should be applied to each aeration basin to ensure good mixing and to keep the biomass in suspension. The aeration equipment must also operate within a prescribed range for safety reasons (surging).

In order to select the manipulation variable and points of application, the type of driving motor, plant size, system configuration, power consumption relationships for keeping the biomass in suspension, final control elements and control objectives must

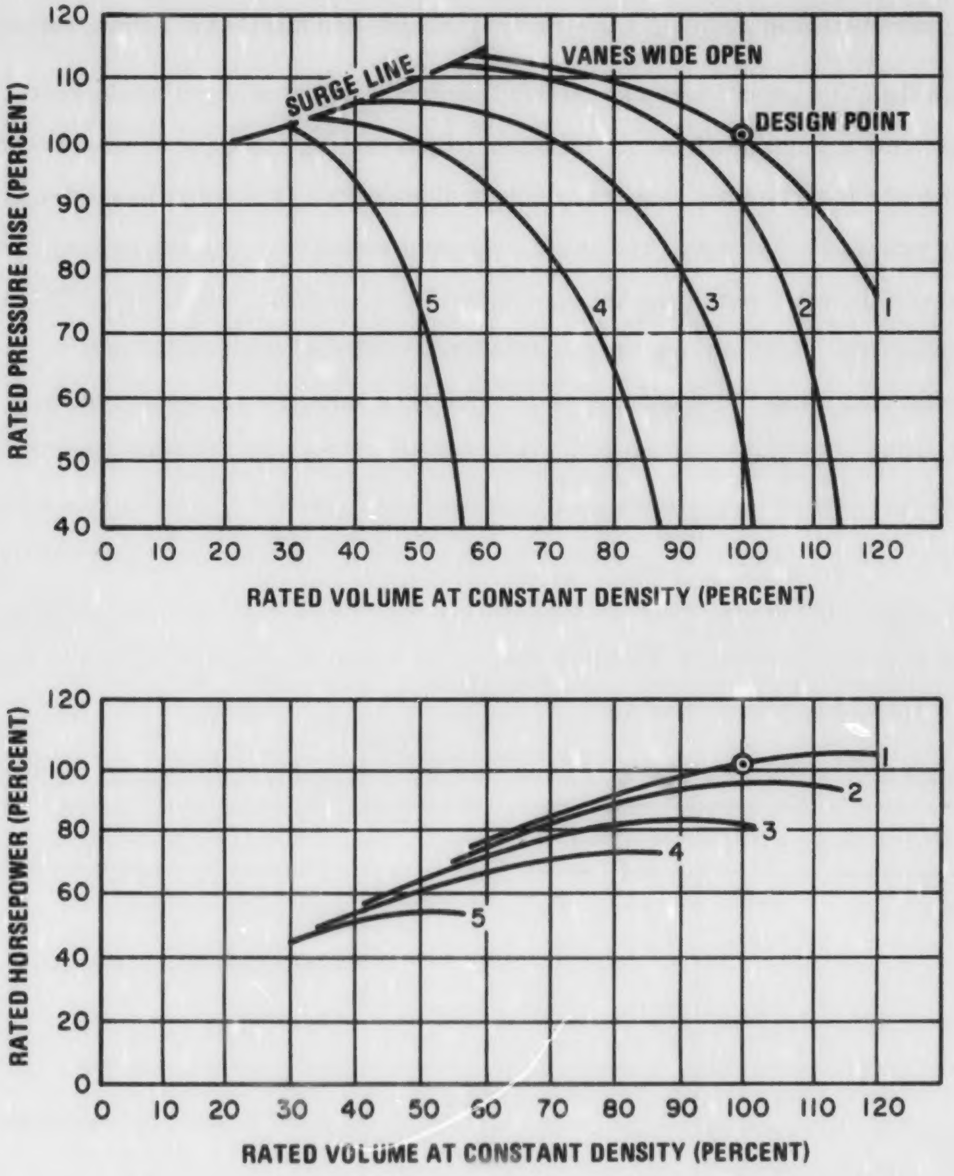

Figure 8. Centrifugal blower performance curves showing effects of inlet guide vanes. (Numerals 1-5 indicate five different settings of vanes.)

be considered. Table 3 contains a list of possible manipulation variables and associated final control elements for mechanical turbine and diffused aeration equipment.

Although the DO control loops can interface directly with final control elements, cascade control loops (both outer and inner loops) are highly recommended because they increase the system's responsiveness and stability. The outer loop adjusts the air flow rate to the aerator. The inner loop compensates for secondary variations and lessens the workload of the DO controller. For example, consider a diffused aeration system where airflow rate at standard conditions is the manipulation variable. Without an inner flow loop, changes in air flow demand could cause excessive demand on its centrifugal blower and surging would result. If its inner loop controlled pressure and sensed temperature, surging could be prevented. For some aeration equipment, such as multi-speed blowers or multi-speed mechanical aerators, the inner cascade loop is nothing more than a simple stepping relay. Other equipment, such as centrifugal blowers, requires rather elaborate inner control loops to throttle airflow rates and prevent surging.

Table 3. POSSIBLE MANIPULATION AND FINAL CONTROL ELEMENTS

Aeration Equipment	Manipulated Variable	Final Control Element
Air diffusers	Airflow rate	Automatic inlet valve
		Automatic inlet valve positioner
		Variable-speed motor
		Multiple-speed motor
		On-off control for multiple blower installations
Submerged aerator (turbine/orifice)	Airflow rate	Same as air diffusers
	Turbine speed	Multiple-speed motor
Surface aerators	Immersion depth	Adjustable weir
		Adjustable blade depth
	Speed	Multiple-speed motor

Proposed Control Strategies

Measuring flow rates, MLVSS, respiration rates, and dissolved oxygen demands would permit virtually perfect control. However, the wastewater characteristics (α, β, and kl_a) continually change in most treatment systems, respiration rates are very difficult to measure accurately, and DO demand measurements are hindered by noisy DO signals.

Since oxygen demand depends on flow rate to some extent, raw sewage flow rate was the earliest proposed control parameter for regulating oxygen transfer rates. The flow-proportioned control system (Figure 9) is an open-loop controller which automatically maintains a constant ratio between the raw wasteflow and airflow at standard conditions. If the strength of the sewage remains constant, a flow-proportional DO

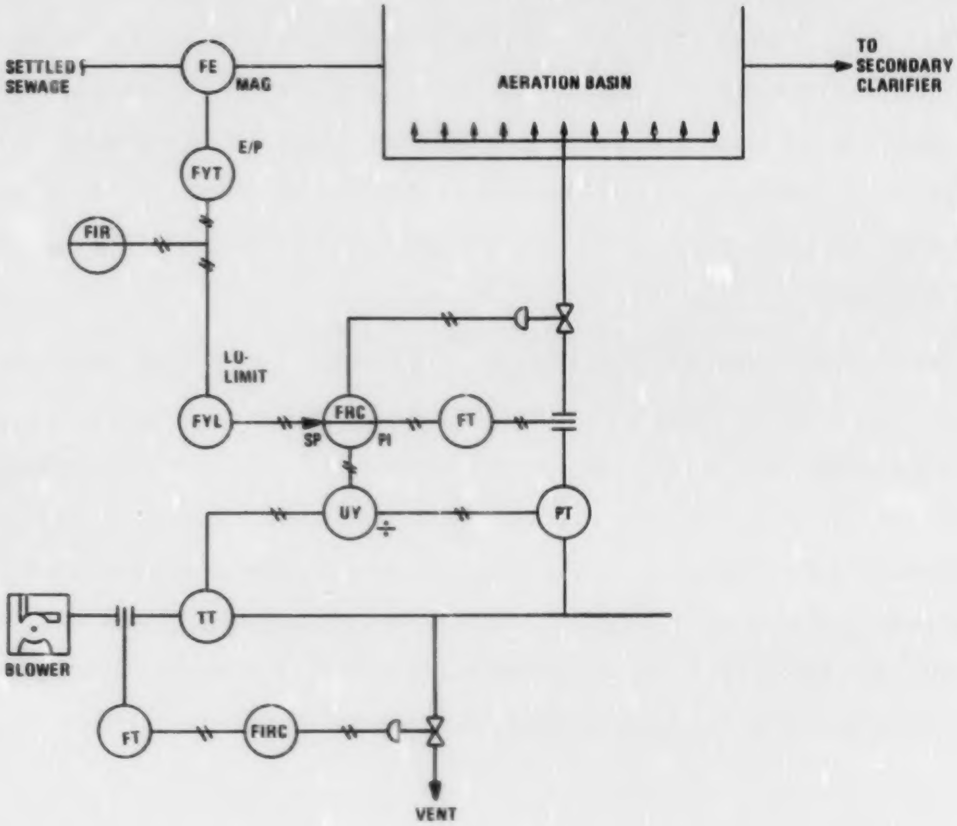
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Figure 9. Flow-proportional DO control for completely mixed activated sludge.

control system will track oxygen demand quite well. However, since most raw sewage strengths change dramatically during a 24-hour period, the ratio must be frequently readjusted, requiring considerable manpower. Instead, most facilities simply choose a high ratio (typically 1.0 to 1.5 feet^3 air/gallon of wastewater) to prevent septic conditions and reduce the adjustment manpower requirements. Aeration power savings, accordingly, are quite small, usually about 10%. Inasmuch as raw sewage flow provides only an indirect estimate of oxygen demand, which cannot distinguish between sanitary wastes, storm water, or infiltrated ground water, flow-proportional DO control is not suitable for most municipal and industrial treatment plants. It is useful only for small industrial plants that have constant strength wastewater.

With the availability of in situ DO probes, feedback control of an aeration basin's dissolved oxygen became practical. Inspection of Equations 10a and 10b shows that airflow rates or power requirements vary directly with the DO operating level. Additionally, Pontryagrin's Minimum Principle says the manipulation variable should be either at the maximum or minimum value when an error exists; otherwise, the steady-state control action is best. Accordingly, optimal control theory requires error data also.

Good DO feedback controllers use high-gain proportional action to give fast responses plus an integral mode to eventually generate suitable steady-state control in the absence of changing oxygen demands, as illustrated in Figure 10. This control strategy, which has many of the desired features indicated by optimal control theory, minimizes secondary disturbances and increases responsiveness by means of a cascaded local loop around the final control element. For diffused air this might be an airflow control loop, or a submergence depth control loop for surface aerators. When a positive error exists (low DO), the feedback controller responds by increasing the airflow rate

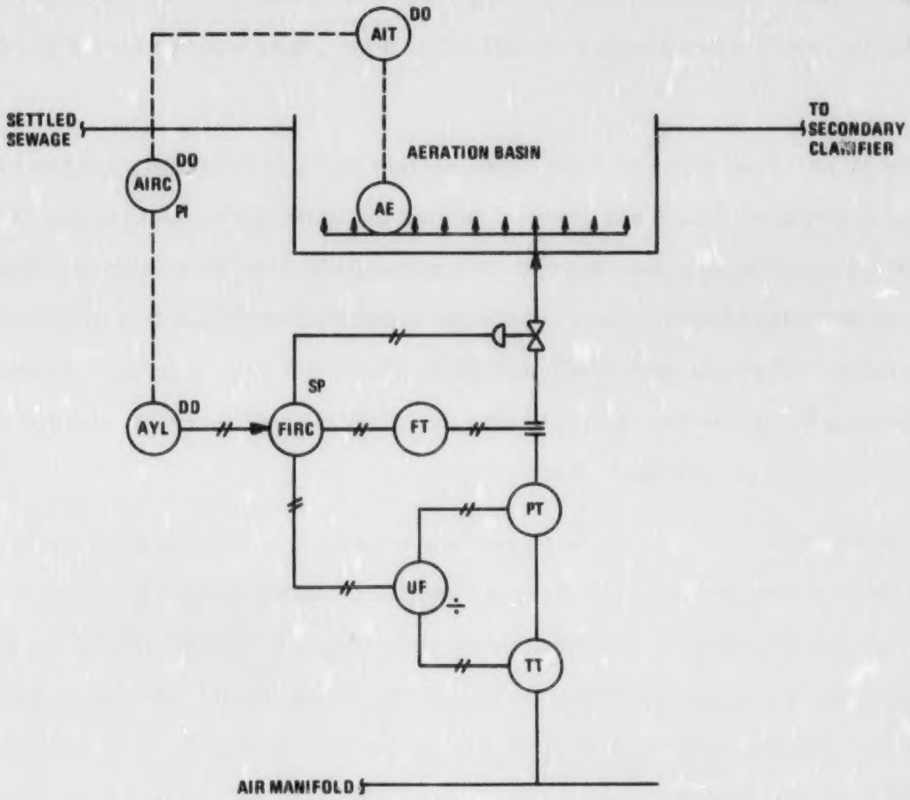
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Figure 10. PI feedback DO control for completely mixed activated sludge.

setpoint (power consumption) and vice versa regardless of the cause of the DO disturbance. Mathematically, the manipulation variable follows Equation 11:

$$U = k_1 (e) + k_2 \int_0^{+} edt \qquad (11)$$

where e equals the desired DO minus the actual DO and k_1, k_2 represent the adjustable controller gains.

If process conditions give rise to sporadic-type DO readings, it may be necessary to filter or smooth the probe reading prior to taking control actions, since feedback DO controllers respond to disturbances, regardless of causes. Other types of feedback controllers, such as two position or differential gap, may also be useful in some cases.

PI feedback DO controllers are the most popular form of DO regulation practiced today in wastewater-treatment plants. During the nationwide plant survey (1), most users of this control scheme reported 25% reductions in aeration energy consumption. Of the six installations observed, four worked satisfactorily and two were considered unacceptable. The principal problem with feedback DO control is the maintenance requirements associated with the DO probe. About 60 manhours are needed annually to inspect, clean, and calibrate each DO probe.

Feedforward control strategies can generate suitable control actions without incurring sizable errors characteristic of sluggish feedback control systems. All feedforward strategies, by definition, are responsive to incoming load conditions rather than waiting for the resultant error before taking corrective action. Based on Equation 10, the forthcoming control strategies neglect the derivative terms and respond to changing influent demands. The feedback control trims the feedforward strategy to compensate for model inaccuracies, measurement errors, and changing ambient conditions.

If a PI controller replaces the respiration rate and derivative terms in Equation 10a, the airflow rate (power consumption) is given by:

$$U = k_1 F\,(S.P.) + k_2 e + k_3 \int e\,dt \qquad (12)$$

where k_1, k_2, k_3 are adjustable controller gains and S. P. = the DO setpoint.

As illustrated in Figure 11, airflow rate regulation (power consumption) is proportional to the influent flow rate and the error in the aeration basin's dissolved oxygen. If the influent flow rate increases, the feedforward loop increases the airflow. When the basin's DO level is low, the feedback loop reinforces this action; otherwise, the

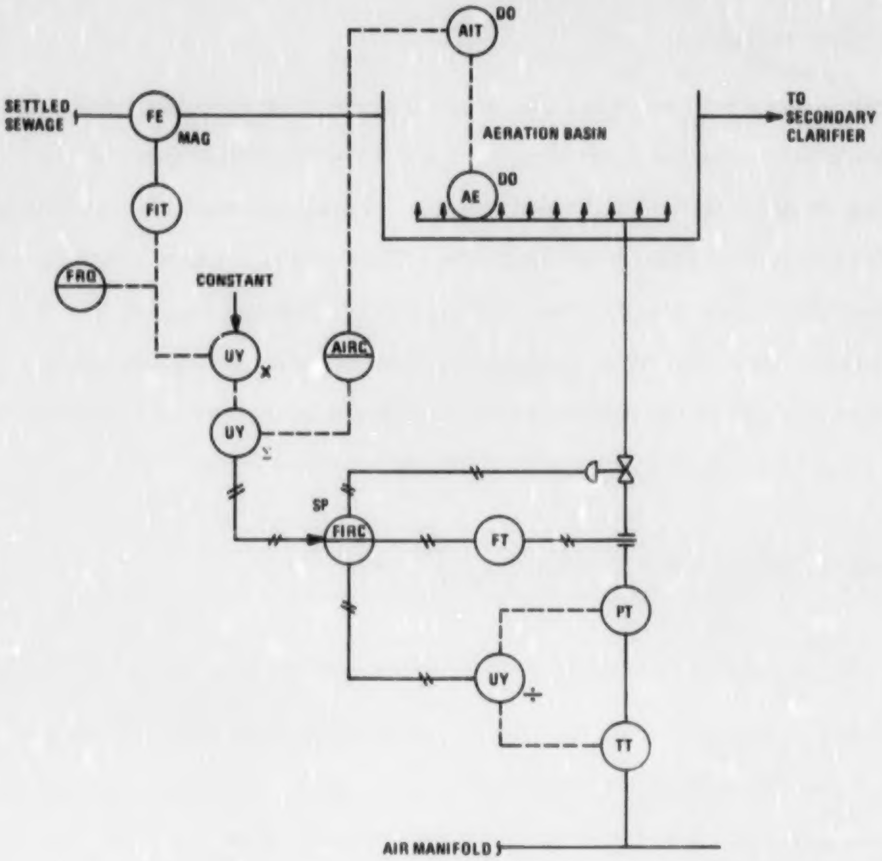
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Figure 11. Flow proportional feedforward DO control with feedback trimming.

feedback control trims the airflow. Similar comments apply to falling influent flow rates. This type of control tends to eliminate diurnal DO oscillations associated with varying flows, although storm-water surges and groundwater infiltration may cause false increases in airflow rates (power consumption).

Because most treatment facilities already have influent flow rate-measuring devices, this control strategy requires the same equipment and maintenance as the feedback DO control loop. In some plants, flow distribution networks may necessitate additional flowmeters, but this appears unlikely. Under EPA sponsorship, evaluations of this control strategy at Palo Alto, California, indicated that power consumption savings were 11%.

Flow and Mass Loading

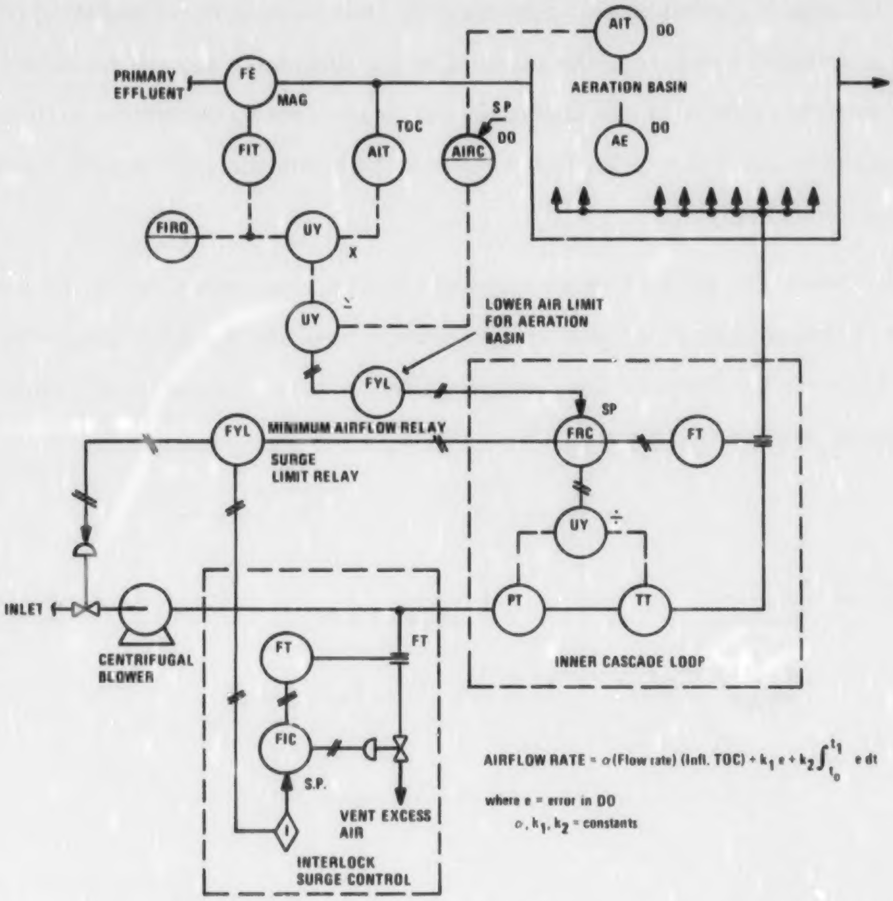
Recent improvements in rapid on-line organic measuring systems make automatic substrate determinations a near-term reality. On-line total organic carbon (TOC), total oxygen demand (TOD), and respirometer (organic demand, OD) instruments are available but require further demonstration. Total organic carbon data in conjunction with influent flow rates can provide estimates of the organic loading ($F * S_i$) on the aeration basin. Since the oxygen demand increases with the organic loading and influent flow rate, aeration power requirements can be determined by rearranging Equation 10b, where PI control replaces the derivative terms:

$$U = k_1 \; F \; (S.P.) + \alpha \; F \; S_i + k_2 e + k_3 \int_0^t edt \qquad (13)$$

where k_1, k_2, k_3 are adjustable controller gains, and S.P. = the DO setpoint.

As illustrated in Figure 12, the feedforward loop measures the total organic load in the influent and regulates the airflow rate accordingly. When the load increases, the feedforward loop calls for more air, but this action is trimmed on the basis of current and past DO levels. If the actual DO is lower than desired, the feedback loop augments the feedforward action or, if the DO level is greater than the setpoint, the feedback loop opposes the forward loop. Although operating experience is necessary to select the appropriate α ratio (ft^3 air: lb of organic), 700 to 1000 ft^3 air/lb of BOD is a typical value for diffused aeration systems. The selected value, however, depends on the operating DO level, mass transfer coefficients, type of treatment, and degree of BOD removal.

Inasmuch as on-line organic monitors are new and still undergoing evaluation in wastewater-treatment plants, this control strategy is untried. Engineering estimates show, however, that 35% power savings are attainable for mass loading feedforward feedback dissolved oxygen control systems.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AIRFLOW RATE = α(Flow rate) (Infl. TOC) + $k_1 e + k_2 \int_{t_0}^{t_1} e\,dt$

where e = error in DO

α, k_1, k_2 = constants

Figure 12. TOC feedforward DO control with feedback trimming.

Respirometers measure the oxygen utilization rate of the aeration basin. If the aeration equipment dissolves oxygen at a rate equal to the utilization rate, the aeration basin's DO level would remain essentially constant. Adding enough additional oxygen to raise the incoming wastewater DO to the operating level ensures an exact balance between oxygenation capacity and demand. Then the DO remains at the desired level. From this concept, on-line respiration rates and flow data can regulate aeration energy consumption according to the following equation, which is derived from Equation 10a:

$$U = k_1 \ F \ (S.P.) + R\left(\frac{V}{F}\right) + k_2 e + k_3 \int edt \qquad (14)$$

Including PI feedback control makes this strategy less dependent on modeling and measurement errors; moreover, the integral mode eliminates permanent offsets. Because respiration rate data are approximately equivalent to derivative action, this strategy anticipates aeration basin DO changes and adjusts the airflow rate (energy consumption) accordingly.

As shown in Figure 13, the airflow rate increases with respiration rates and may increase or decrease with flow rates in the forward loop; the feedback loop either reinforces or trims the forward loop control action in such a manner as to reflect the current state of the aeration basin.

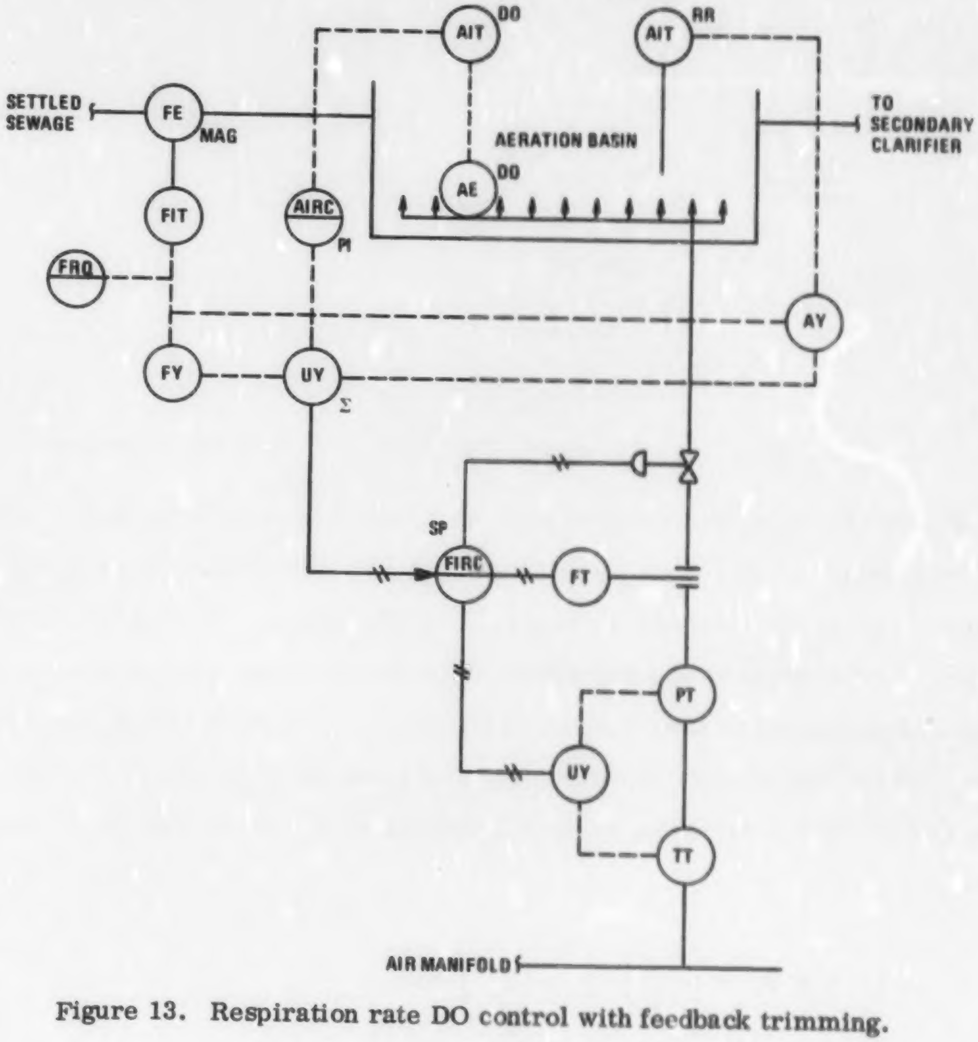
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Figure 13. Respiration rate DO control with feedback trimming.

The oxygen demand equation (Equation 10) can be readily solved for the respiration rate, R, if the airflow rate, residual DO, and transfer coefficient are known; moreover, derivatives of the oxygen concentration can be numerically estimated. Practical considerations, however, dictate using a small on-line computer or time-sharing with a larger facility for estimating the respiration rate. Implementation of calculated respiration rate control must follow the difference equation form of Equation 10a:

$$U_{(j+1)} = U_j - \frac{Co_j + 1 - Co_j}{\Delta t} + k_1 \,(SP - Co_j + 1) + k_z \sum_{N=1}^{j+1} (SP - C_N)\,\Delta t \qquad (15)$$

where

$$SP - Co_{j+1} + e_{j+1}$$

$$\sum_{N=1}^{j+1} (SP - C_N) \quad t = \int_{+0}^{t} edt$$

$$e\ (0) = 0$$

The PI control mode was added to ensure convergence to the setpoint. Although influent flow rate, airflow rates, and transfer coefficients are needed to calculate the respiration rate, when the expression for this is substituted into Equation 10a, these terms cancel. Since Equation 15 is the digital approximation for three-mode proportional integral derivative (PID) control, calculating the respiration rate is equivalent to approximating the aeration basin's first derivative, and the resulting control strategy becomes the classical feedback PID controller. Unlike the direct respirometer measure technique, this method adds very little information about system behavior and is not a true feedforward control strategy. The benefits and maintenance requirements are virtually the same as PI feedback DO control.

Conventional Activated Sludge—Plug Flow Configuration

Many secondary wastewater-treatment plants are built using a long, narrow, multipass aeration basin which has become known as the plug flow configuration. In actuality, the backmixing caused by the rolling action of aeration units makes this type vessel behave like several completely mixed reactors in series. Consequently, completely mixed analyses extend to the plug flow rather directly. The key question becomes: how many tanks in series are needed to represent the plug flow basin? Previous investigators (6) have used from two to ten vessels to represent the conventional activated sludge process. Since the oxygen demand changes along the length of the operation basin, as shown in Figure 14, finding the optimum number and location of DO probes is a difficult task. At the head or inlet of the aeration basin where the return biomass and primary effluent are mixed, the organic concentration (and hence the oxygen demand) is highest. It is important to supply enough DO so as not to inhibit stabilization; a DO of approximately 0.5 mg/l should be sufficient. As biological stabilization proceeds along the length of the aeration basin, the oxygen demand

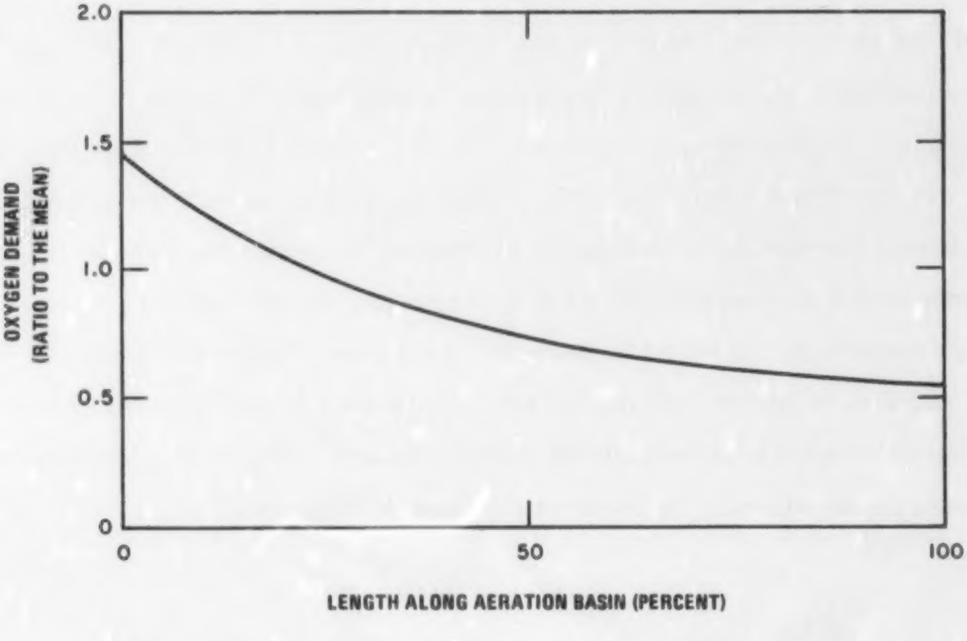


Figure 14. Variation in oxygen demand with basin length.

diminishes. If properly designed, the oxygen demand of the outlet end is small and a DO concentration between 1.5 and 2.0 mg/l is desired.

The selection of a control strategy and DO probe placement depends heavily upon the type and flexibility of the aeration system. When a uniform aeration intensity is used, the oxygen profile is virtually fixed, and only the level (rather than the shape) can be altered via automatic control. In the influent zone, oxygen demand is high and the resultant DO level is low; farther down the aeration tank, the oxygen demand is lower and the resultant DO is high. If a separate aeration source which can be regulated exists for each pass, then more satisfactory control is possible because it can regulate power consumption and the DO level for each pass. In the following paragraphs some notes on the application of automated DO control to plug flow aeration are provided.

Plug flow aeration basins are designed as single- or multiple-pass units, and the oxygen transfer equipment is either uniformly applied or divided into separately adjustable sections. Single fixed output systems have little flexibility and do not lend themselves to automatic regulation. Although aeration theory maintains that the head section has the highest oxygen demand and consequently should be the best location for DO probes, alternate probe placement should be planned so that the operator by trial and error techniques can find the optimum probe location. Recommended control strategies for this type (listed in Table 4) are simple to tune and keep operational.

Separately adjustable aeration systems can be considered similar to several completely mixed systems in series. Clearly the probes for the respective control systems should be placed in their relevant zones of influence. Here, correct probe placement is important, but not as crucial as with uniform aeration. Control strategies (listed in Table 5) range from simple techniques requiring no special equipment to sophisticated control systems needing sophisticated equipment and maintenance. Potential power savings usually increase with the complexity of the control system. The best strategy depends on many factors, such as plant size, power costs, and maintenance labor.

Table 4. SUMMARY OF DO CONTROL STRATEGIES FOR PLUG FLOW AERATION BASINS WITH UNIFORM OXYGEN TRANSFER

Description	Aeration Power Savings	Advantages	Disadvantages	Recommendations
Flow Proportional	5%	Simple, straightforward, easy to maintain	Does not respond to change in organic	Suitable for small plants with uniform oxygen
PI feedback DO control	10%	Aeration power responds to aeration basin DO level	Controls DO at a single point only along the aeration basin	Good controller for medium-size and large plants

Economic analyses are applied as decision making aids in a subsequent section, where specific control strategies are recommended.

One method of automatic DO control is flow ratio control, but adjusting aeration equipment capacity in proportion to influent flow rate provides only marginal savings since the wastewater composition undergoes considerable fluctuation in a 24-hour period. Moreover, such a control system is prone to enormous errors during storm events and when infiltration occurs. As in completely mixed systems, only about 10% power savings can be expected and poor DO regulation is customary with this type of control system.

In the case of feedback control, in situ DO probes, located in each zone of influence coupled with PI control, pace aeration capacity in accordance with local DO levels. For example, a four-pass diffused aeration system with each pass capable of separate regulation can keep the dissolved oxygen level of each pass at the prescribed level. In concept, each pass is treated as a separate completely mixed vessel. Sufficient flexibility must be provided for placing the probe anywhere in the aeration pass. Plant operators, by virtue of trial and error probe locations, will find the optimum probe placement for each pass. When surface aerators with speed or individual submergence controls are installed, the DO probe should be located in each aerator's zone of influence.

Table 5. SUMMARY OF DO CONTROL STRATEGIES FOR PLUG FLOW BASINS WITH SEPARATELY ADJUSTABLE OXYGEN TRANSFER DEVICES

Description	Aeration Power Savings	Advantages	Disadvantages	Recommendations
Flow proportional control	5%	Simple, straightforward, easy to maintain	Does not respond to changes in organic strength	Suitable for small uniform aeration intensity plants
PI feedback DO control	15%	Aeration power responds directly to actual basin DO level	May tend to oscillate due to flow variations	Good controller for moderate sized, uniform, and separately adjustable aeration systems
Flow proportional with feedback trimming	20%	Reacts to flow rate variations and adjusts the aeration intensity profile accordingly	May be difficult to tune this controller. High infiltration rates or storm events may lead to poor control	Applicable only for large, separately adjustable aeration systems
Organic loading with feedback trimming	25%	Responds to flow and strength variations in the influent stream	This untried strategy requires an on-line organic monitoring system which may be difficult to maintain	Useful in large, highly automated plants, especially with F/M control, that have separately adjustable aeration systems
Power forward with feedback control	22%	Responds to flow and variations in the first aeration section; consequently, DO demand changes are anticipated for the later sections	May not be any more beneficial than PI controller	Suitable for moderate sized, separately adjustable aeration systems

PI feedback DO control systems of separately adjustable aeration systems can reduce aeration cost by 25%; each control loop requires 60 manhours of maintenance per year (the same as completely mixed feedback loops).

A uniform aeration system has less flexibility since only the total power can be regulated rather than a series of individual aerators. Automatic regulation of this type of system consequently can only raise or lower the entire DO profile. Inasmuch as the plug flow is equivalent to the transportation lag, the DO probe should be located in the leading section of the aeration basin; the changes in the oxygen demand are also most severe in this section. However, the plant operators must still be able to move the DO probe so as to find the best location. Some design engineers recommend monitoring the middle and end sections DO, but rarely is this information used for control purposes. The PI feedback control of uniform aeration systems reduces power costs by approximately 20%, and 60 manhours of maintenance per system are required.

Unlike completely mixed aeration basins which equalize the influent load, plug flow basins keep the inflow segregated. Accordingly, the head section reacts rapidly to changing influent oxygen demands and can be used to forewarn the later sections. In other words, a feedforward control strategy in which the control information flow follows the hydraulic flow pattern is well suited. Since plug flow reactors have spatial as well as temporal variations in concentration and reaction rates, their behavior must be described by partial differential equations. A material balance around the differential section of a plug flow basin yields Equation 16, whose solution describes the time and distance relationships to dissolved oxygen:

$$-Q \frac{\partial C_o}{\partial L} + r + kl_a (C - C_s) U (L) = \frac{\partial C_o}{\partial t} \qquad (16)$$

where:

Q = space velocity (flow rate/normal area)

L = length or distance from inlet

t = time

r = respiration rate

At steady state:

$$\frac{\partial C_o}{\partial t} = 0$$

$$\frac{\partial C_o}{\partial L} = -\frac{r}{Q} + \frac{kl_a}{Q} (C_o - C_s) U (L)$$

If U (L) = constant, then:

$$\frac{C_o}{L} = -\frac{r (L)}{Q} + \frac{kl_a}{Q} (C_o - C_s) U$$

It can be shown that $\partial C_o / \partial L > 0$ under these circumstances and a DO gradient exists in the aeration basin.

Under constant loading conditions and using second-order kinetics (e.g., $r = kS\overline{X}$), then:

$$U (L) + \alpha k S_{in} e^{-k/Q L \overline{X}} - \beta kd \overline{X} \qquad (17)$$

where:

$\overline{X}$ = average MLVSS

kd = endogenous respiration constant

k = second order rate constant

Tapered aeration attempts to match the oxygen transfer to the oxygen demand according to a relationship similar to Equation 17. It is informative to notice that increases in influent concentration (S_{in}) raise the entire aeration requirement, but the distribution profile remains virtually the same; on the other hand, flow rate changes alter the shape of the air distribution network.

In summary, uniform aeration systems are difficult to control and the DO concentration must vary throughout the reactor. The head section always has a lower DO value than the last sections. Lack of flexibility precludes the use of any elaborate feedforward control scheme. Flow ratio, feedback DO, or combinations of these two should provide meaningful control of uniform aeration systems. The saving attainable with these control systems is somewhat less than in the completely mixed cases since only single point control is possible.

For separately controllable aeration systems the following advanced but not yet demonstrated control strategies will improve DO profile regulations. Flow and feedback DO control will be discussed first. Equation 17 shows that the airflow (power consumption) requirements at any given location change exponentially with respect to hydraulic flow rate perturbations. Adding PI feedback control ensures elimination of measurement and modeling errors:

$$U = k_1 \exp(-C_1/F) + k_2 e + k_3 \int e\,dt \tag{18}$$

where:

$C_1 = kA\overline{X}L$, with A = cross-sectional area normal to flow direction.

If the aeration vessel has four separate control loops, tuning involves selecting 12 gains and four constants. Obviously, this controller is too complicated for other than large plants.

Engineering estimates indicate 20% power savings with this control strategy. Approximately 60 manhours per year of maintenance is anticipated for each control loop. Since most plants that practice plug flow aeration use either two or four passes, the total maintenance demand is the product of the number of loops (passes) multiplied by the maintenance requirements per loop.

Another approach is to use feedforward substrate and flow with feedback DO trimming for control. Increased influent substrate concentrations (S_{in}) result in a proportional increase in aeration intensity along the entire length of the aeration basin, as seen from Equation 18. Flowrate increases, as discussed in the above control scheme, necessitate an alteration of the aeration profile. Substrate measurements via TOC, COD, TOD or respirometry and flowrate information are combined according to Equation 17 into the following control, where PI feedback control is added to offset measurement and model errors:

$$U = k_1 S_{in} \exp(-C_1/F) + k_2 e + k_3 \int e\,dt \tag{19}$$

This control strategy adjusts the airflow (power consumption) for each pass or loop in proportion to the influent substrate and exponentially with respect to the flowrate. Rather than utilizing a substrate monitoring instrument, it may be possible to monitor the power consumption or airflow rate for the first pass while under the control of PI DO feedback regulation and feed it forward to the remaining aeration loops. If U_i denotes airflow rate or power consumption for the jth section, this control strategy functions according to the following control law:

$$U_j = k_{1j} U_{j-1} \exp(-C/F) + k_2 e_j + k_3 \int e_j\,dt$$

$$U_1 = k_{1_1} c + k_{2_1} \int e\,dt$$

Engineering estimates show that this untried control strategy should reduce aeration power by 25% for the substrate monitoring version, and by 20% for the feedforward power consumption technique.

Even for completely mixed aeration systems proper DO probe location is essential for all but the flow ratio control strategy since the control system acts on local DO concentration sensed by the in situ probe. Accordingly the probe should be located in a representative region of the aeration vessel. (The stagnant areas with low local DO levels are unrepresentative regions and should be avoided.) Inasmuch as completely mixed systems have large, well-mixed areas with uniform DO concentrations, finding a representative region is usually easy.

Two methods are currently used for DO sensing: in situ and remote monitoring. In situ methods are preferred because the time lags are essentially zero. All in situ arrangements should have enough flexibility to allow operators to easily move the DO probes to several locations in the aeration basin. If the DO probe is remotely located from the aeration basin, the elapsed time and subsequent oxygen depletion in transporting the mixed liquor sample from the basin to the probe must be included in the setpoint selection and control system tuning.

Although the DO control loops can interface directly with final control elements, cascade control loops are highly recommended because they increase the system's responsiveness and yet are stable. For some aeration equipment, such as multi-speed blowers or multi-speed mechanical aerators, the inner cascade loop is nothing more than a simple stepping relay. Other equipment, such as centrifugal blowers, requires rather elaborate inner control loops to throttle airflow rates and prevent surging.

All of the proposed control systems require varying amounts of maintenance. The primary sensors, DO probes, organic analyzers, etc., must be kept calibrated to provide accurate data; moreover, controllers must be kept properly tuned. Table 6 summarizes the operational and maintenance characteristics of the proposed DO control strategies for completely mixed aeration basins.

Since aeration basins exhibit imperfect mixing at the microscale level and available DO probes exhibit some noise, instantaneous DO probe readings may not accurately represent the true or average DO concentration in that region. Field experiences support those comments. For example, in completely mixed aeration systems, turbulent mixing conditions give rise to macro-fluid elements moving about the vessel; accordingly, the DO fluctuates as these macro-elements pass across the DO probe surface. The same phenomenon also occurs at a given cross-section of a plug flow aeration basin (7). Since macro-mixing under turbulent conditions is a random-type process and the noise associated with DO probes is also random, reasonably the sporadic DO readings may be viewed as random events and the DO signal as the sum of the true value and some random variable. Moreover, the available data indicate that the randomness can be considered as white (random) noise with a zero mean.

The anticipated difficulty emerges when estimating the true DO value from the probe signal. Estimators range from simple averaging delay and hold circuits to Kalman filters and to nonlinear regression analyses. Choosing the best filter (estimation procedure) is difficult. However, in line with the preferred philosophy of keeping a control system as simple and easy to maintain as possible, the straightforward estimators such as multipoint or time averaging, short-time rejection, delay and hold, and low-pass filters are adequate for aeration basin DO control.

The quality of automatic regulation obtained from the proposed control systems depends heavily on the adjustments made to their control modes. Without proper tuning, all of the feedback controllers and some of the feedforward strategies will perform poorly. It should be noted that the degree of difficulty increases with the number of control modes. All the feedback controllers should be adjusted via the Ziegler-Nichols method of ultimate sensitivity, with subsequent field experience providing added refinements; the feedforward controllers should be adjusted by trial and error.

Table 6. SUMMARY OF AUTOMATIC DO CONTROL FOR COMPLETELY MIXED AERATION BASINS

Description	Aeration Power Savings	Confidence Level	Advantages	Disadvantages	Recommendations
Flow proportional	10%	High	Simple, straightforward control strategy.	Frequently yields poor DO regulation, especially with storm water or ground water infiltration.	Useful only in small plants (less than 1 mgd), or for constant organic strength wastewaters.
PI feedback	25%	High	Control based on actual DO values. Most commonly practiced automatic DO control. Easy-to-tune controller.	Slow response to upsets; also may tend to oscillate. Many times, a filter is required to reject sporadic DO signals.	Suitable for most wastewater-treatment facilities.
Flow proportional with DO feedback trimming	30%	Medium	Fast response to flow rate changes. Readily adaptable to most plants.	Moderately difficult to tune. High infiltration or storm events may lead to poor control. Not any more sensitive to oxygen demand changes than PI feedback controller.	Very useful in moderately large plants with dry-weather flow and relatively constant organic strength influent.
Organic loading via TOC with feedback trimming	35%	Low	Responds to influent load changes (flow and organic) before they upset the aeration basin.	This untried control strategy requires an on-line organic monitoring instrument which may prove difficult to maintain.	Suitable only for large, highly automated plant. Should also be cascaded with F/M control.
Calculated respiration rates with feedback trimming	30%	Medium	Uses aeration power levels at constant DO values to calculate respiration rates. These measuring devices are readily available and well behaved.	Requires a process computer. Control may be no better than simple PID control.	Useful in large plants with real-time computer time available.
Aeration basin respiration rates with feedback trimming	35%	Low	Since this control reacts to respiration rate change, it anticipates DO variations and adjusts aeration power accordingly.	High maintenance demands and poor reliability of on-line respirometers are a serious drawback to this strategy.	

FOOD-TO-MICROORGANISM CONTROL

Introduction

With proper operation the activated sludge process can achieve 90% to 95% BOD reductions. For most municipal wastewater facilities, this means production of an effluent BOD of about 20 mg/l or less. To operate an activated sludge process efficiently, it is necessary to control physical and biological parameters such as temperature, dissolved oxygen, hydraulic residence time, food-to-microorganism ratio (F/M), and mixed liquor suspended solids. This section addresses control of one of the most important biological parameters—the F/M.

In the activated sludge process, wastewater is contacted with microorganisms under aerobic conditions whereby the microorganisms use the biodegradable pollutants as a food source and purify the wastewater. It is essential for the microorganisms to form a satisfactory floc so that they can be effectively separated; a fraction of the settled biological solids is then recycled and the remaining mass is wasted.

It is generally accepted that BOD removal efficiency is dependent on the F/M. If this ratio is too high, the bacteria undergo exponential growth and form dispersed floc that does not settle and, as a consequence, the BOD removal is unsatisfactory. On the other hand, for low F/M, unoxidized fragments of the floc remain in suspension and cause a turbid effluent.

By means of the material balances on the substrate (BOD) and the biomass around an activated sludge system (illustrated in Figure 15), it can be shown that the mean cell residence time is related directly to the F/M:

$$\frac{1}{\theta_C} = YU - kd$$

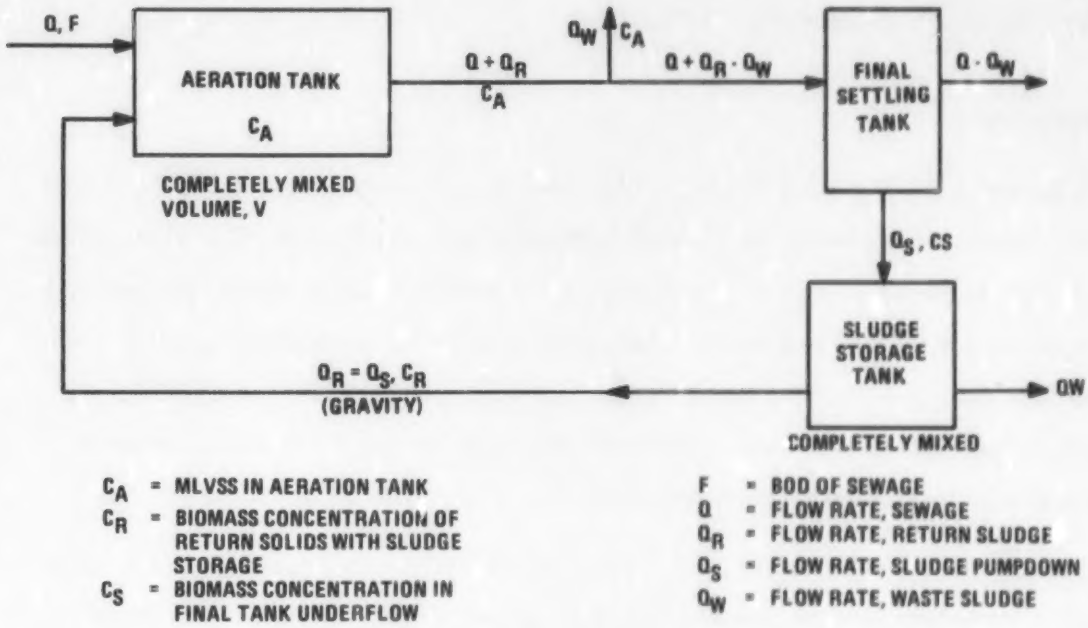

Q, F
AERATION TANK
C_A
COMPLETELY MIXED VOLUME, V
$Q + Q_R$
C_A
Q_W
C_A
$Q + Q_R - Q_W$
FINAL SETTLING TANK
$Q - Q_W$
Q_S, CS
SLUDGE STORAGE TANK
COMPLETELY MIXED
QW
$Q_R = Q_S, C_R$
(GRAVITY)
C_A = MLVSS IN AERATION TANK
C_R = BIOMASS CONCENTRATION OF RETURN SOLIDS WITH SLUDGE STORAGE
C_S = BIOMASS CONCENTRATION IN FINAL TANK UNDERFLOW
F = BOD OF SEWAGE
Q = FLOW RATE, SEWAGE
Q_R = FLOW RATE, RETURN SLUDGE
Q_S = FLOW RATE, SLUDGE PUMPDOWN
Q_W = FLOW RATE, WASTE SLUDGE

Figure 15. Activated sludge system.

where:

Y = growth-yield coefficient mass of microorganisms/mass of substrate utilized

kd = microorganism decay coefficient, time^{-1}

U = F/M

θ_C = mean cell residence time

Organic loading in most plants changes considerably during the day. For example, the BOD loading for Baltimore (as reported by Keefer) (8) varies from 37 percent to 166 percent of the daily average in about an 8-hour period. (This can be seen in Figure 16.) Under such varying loads, it is not possible to maintain a constant F/M ratio by adjusting the sludge wasting rate. This is because the growth rate of biomass is much lower than the rate of increase in BOD loading. Under the loadings shown on Figure 16, if the F/M ratio were at its desired value at 0800 it would be much greater than the desired value at 1600, even with no sludge wasting during that period.

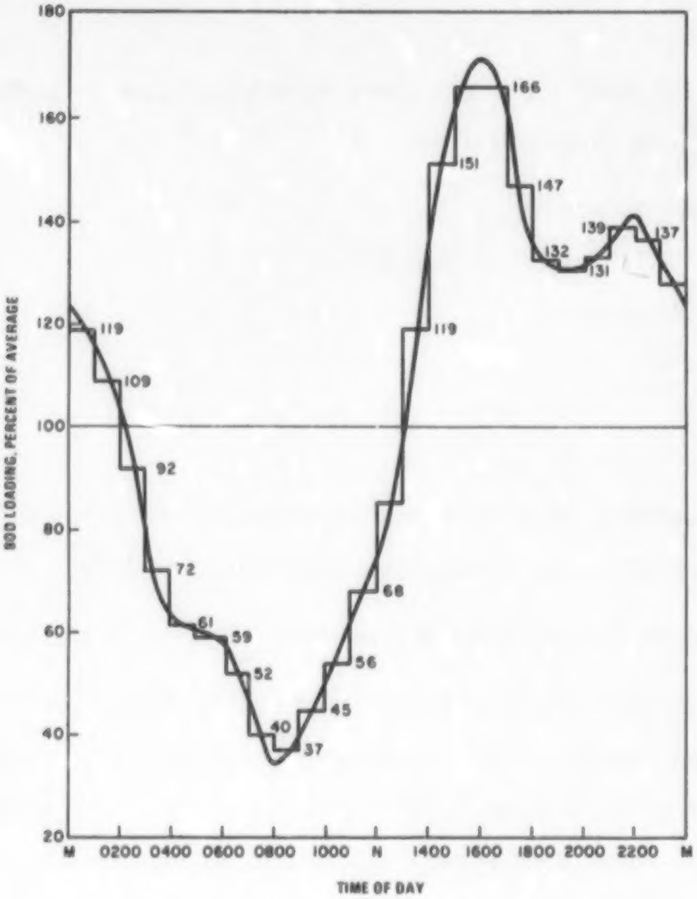


Figure 16. Variation in BOD load in Baltimore.

The diurnal spread in the F/M ratio can be greatly reduced by the provision of sludge storage in the return sludge line, as shown by the flow diagram, Figure 15.

F/M Control Algorithm

A control procedure would involve continuous measurement of the flows and concentrations. F (food) would be measured as TOC, COD or some other parameter which could be correlated with BOD. The manipulated variables would be Q_R and Q_S. Q_W could be automatically provided for as an overflow from the sludge storage tank. The control algorithm, and its derivation, are given in the appendix.

Example

Table 7 gives the results of computations made to simulate the control procedure under the following plant conditions:

Average Q = 70 mgd

Average F = 144 mg/l = 1200 lb/MG

V = 30 million gallons

G = 3.5 days

Y = 0.5

K_D = 0.05

$C_S = C_R$ assumed to be constant at 10,000 mg/l = 83,400 lb/MG. (This would correspond to an average sludge volume Index of 100.)

The diurnal variations in Q and F are in proportion to the values in Figure 16.

The results indicate that it would be necessary to stop the sludge return during the period from midnight to noon. This is because the reduction in C_A brought about by the natural washout of solids would be less than the reduction in C_A necessary to maintain a constant F/M ratio. Thus it would be impossible to maintain a constant ratio during this period. The sludge in the underflow from the final tank would be stored up during this period and would be available to supply the needs for sludge return during the period from noon to midnight.

It is noted that the quantity of sludge storage varied from zero to 4.52 million gallons, assuming no sludge wasting. Thus, a storage tank capacity of about 5 million gallons (one-sixth of the aeration tank volume) would have sufficed. If the settling quality of the sludge were poor (sludge volume index greater than 100) Q_S, Q_R and maximum sludge storage volumes would all be greater than shown.

Analysis

The control strategy requires real-time BOD and mixed liquor volatile suspended solids data, which are difficult parameters to measure accurately. Moreover, a

Table 7. RESULTS OF COMPUTATIONS OF SIMULATED CONTROL PROCEDURE UNDER SPECIFIED PLANT CONDITIONS

T	Q	F	C_{AO}	C_R	C_{AD}	Q_R	β	$e^{-\beta t}$	θ	C_{At}	$\bar{C}_A$	Q_S	V_{SS}
24	70	1626	16,273	83,400	13,279	0	2.3833	0.88865	795.94	14,548	15,410	12.9	1.32
01	68	1543	14,548		12,400	0	2.31666	0.90799	754.85	13,279	13,914	11.3	1.79
02	65	1460	13,279		11,072	0	2.21666	0.91178	713.53	12,170	12,725	9.9	2.20
03	61	1251	12,170		8803	0	2.08333	0.91685	610.49	11,208	11,689	8.6	2.56
04	58	1001	11,208		6773	0	1.98333	0.92069	487.88	10,356	10,782	7.5	2.88
05	54	1043	10,356		6759	0	1.8500	0.92543	507.41	9621	9988	6.5	3.14
06	50	1084	9621		6323	0	1.71666	0.93098	526.21	8999	9310	5.6	3.37
07	56	792	8999		5174	0	1.91666	0.92325	385.66	8337	8668	5.8	3.62
08	62	500	8337		3617	0	2.11666	0.91558	244.09	7654	7996	5.9	3.86
09	65	417	7654		3162	0	2.21666	0.91177	203.80	6997	7326	5.7	4.10
10	67	334	6997		2611	0	2.28333	0.90924	163.83	6377	6687	5.4	4.32
11	76	584	6377		5178	0	2.58333	0.89796	288.21	5755	6066	5.5	4.56
12	85	792	5755		7855	27.0	3.7833	0.85416	20,154	7855	6805	9.1	3.81
13	88	1126	7855		11,560	47.4	4.5667	0.82673	29,238	11,560	9708	15.8	2.39
14	90	1460	11,560		15,330	55.6	4.9033	0.81533	31,975	15,330	13,445	23.5	1.05
15	90	1626	15,330		17,073	39.7	4.3733	0.83342	25,793	17,073	16,201	25.2	0.45
16	90	1751	17,073		18,390	38.0	4.3166	0.83539	25,074	18,390	17,732	27.2	0.00
17	89	1585	18,390		16,453	1.7	3.0733	0.87981	2274	16,453	17,422	18.9	0.72
18	88	1418	16,453		14,556	0.0	2.9833	0.88311	697	14,611	15,332	16.2	1.38
19	87	1418	14,611		14,393	15.2	3.4566	0.86586	12,987	14,393	14,502	12.5	1.48
20	86	1418	14,393		14,227	15.5	3.4333	0.86671	13,148	14,227	14,310	17.4	1.56
21	86	1485	14,227		14,895	24.5	3.7333	0.85594	18,864	14,895	14,561	19.3	1.54
22	85	1547	14,895		15,341	22.9	3.6466	0.85904	18,059	15,341	15,118	19.6	1.50
23	88	1585	15,341		16,273	30.0	3.9833	0.84707	21,435	16,273	15,807	11.0	0.71
Ave	75	1200											

mini- or micro-computer is necessary to implement on-line F/M control. Although all of the above-mentioned equipment and analyzers are commercially available, no wastewater plant currently uses on-line automatic F/M control, but several municipal plants are planning to control the F/M ratio soon. At the present time, however, the benefits of automatic F/M control remain unclear and controversial. Instantaneous F/M control is not needed except where shock-unabsorbable loads occur. Instead, most secondary treatment plants should maintain control at an average F/M with a variation of ±25% allowed.

TRICKLING FILTERS

The trickling filter is the earliest and currently the most widely employed method of secondary treatment. The wastewater stream is distributed over the top of the filter packing by a spray arm that is either hydraulically or electrically driven. The distributed wastewater then flows down by gravity over the media, which has a fixed biological film attached to its surface. Aerobic, facultative, and anaerobic bacteria present in the biological film purify the wastewater by metabolizing the organic contaminants. As in the activated sludge process, removal efficiency depends upon the DO content, nutrient level, pH, temperature, and organic loading/microorganism ratio. BOD removal efficiencies range from 80% to 90% for well-designed and well-operated systems. After passing through the filter, the effluent is collected and sent to a secondary clarifier. Trickling filters are usually classified as low rate, high rate, or super rate, according to their loading rates.

Most of the design principles and operating practices of trickling filters were empirically derived by a group of engineers and scientists acting under National Research Council (NRC) sponsorship. The NRC found that trickling filter efficiency varied inversely with load, and directly with filter surface area and the number of passes of waste through the filter. Equation 20 summarizes the factors that affect operating efficiency:

$$E = \frac{100}{1 + 0.0085\sqrt{W/VF}} \qquad (20)$$

where:

E = efficiency of removal of 5-day BOD

W = weight, 5-day BOD applied per 24 hours

V = filter volume acre per foot

F = number of effective passages through filter

For a single-stage filter with no circulation, F is equal to unity. However, when recirculation is used, F changes as follows:

$$F = \frac{1 + R}{(1 + 0.1R)^2} \qquad (21)$$

where:

R = recirculation ratio

Substituting this relationship into Equation 20 shows that BOD removal efficiency increases with higher recirculation ratios.

As previously mentioned, trickling filter performance is a function of the DO, nutrient levels, and recirculation rates. Recirculation keeps the media moist when flow is low, improves distribution, retards the entry and egress of filter flies, and maintains a sufficient hydraulic load to prevent clogging.

It is desirable to operate trickling filters at maximum BOD removal efficiency and with a minimum amount of power consumption. Equation 20 shows that higher recirculation rates (up to flooding) will increase the BOD removal efficiency. Note, however, that the effectiveness decreases at high recirculation rates. Also, high recirculation rates consume large amounts of electrical energy in the recycle pumps.

Recirculation rates may be held constant, kept above a certain minimum value, or adjusted proportionately to influent flow rate. If constant recirculation is to be used, then process control is unnecessary because process requirements will be on an average basis. However, this mode of operation is unresponsive to changing flow rate.

When the total flow to the trickling filter is to be kept above some minimum value, the total flow must be monitored so that, if the flow falls below a preset minimum, the recirculation pump will automatically turn on; otherwise, the recirculation pump will turn off. Minimum flow control systems prevent the trickling filter from starving during low-flow periods such as nights, weekends, and holidays.

Flow proportional recirculation control, as illustrated in Figure 17, responds to the flow rate variations, which are usually the most significant disturbances encountered by trickling filters. A flow meter monitors the raw sewage flow and forwards this

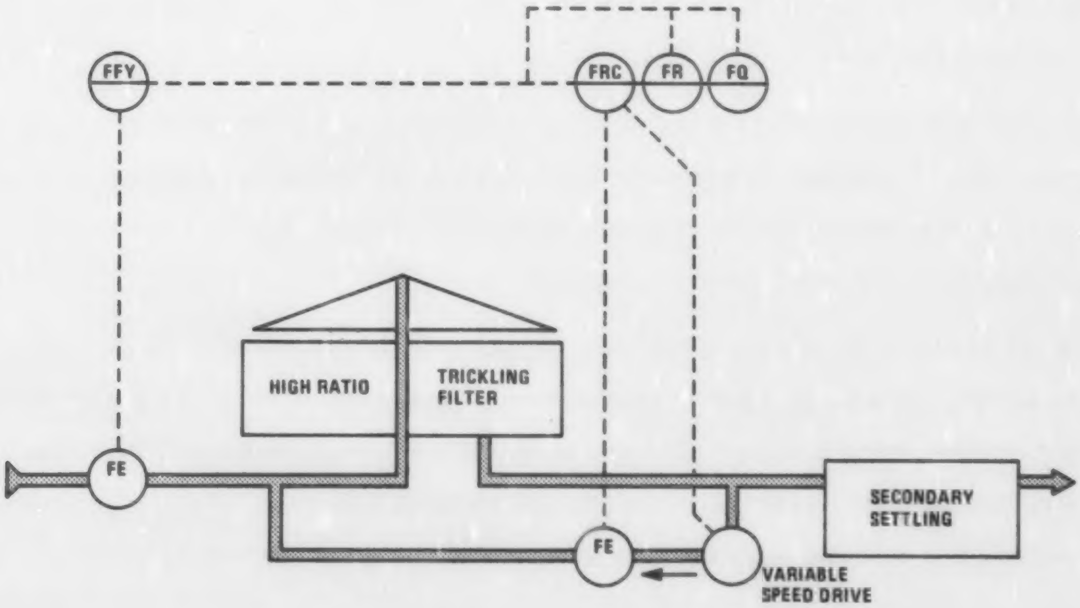


Figure 17. Flow proportional recirculation control.

information to a ratio controller that sets the desired flow rate for the inner recirculation flow control loop. A butterfly valve or variable-speed pump may be used to vary the recirculated flow in proportion to the primary effluent. Flow-proportioned control is relatively simple to implement and provides the following advantages:

- Maximum efficiency of maximum flows without hydraulic overloading
- Minimum necessary recirculated rates to prevent "starvation" at low flows.

The large capacity that is inherent in trickling filters with recirculation makes the systems virtually immune to shock loads. In fact, trickling filters are frequently used as "roughing" filters to protect downstream activated sludge processes from shock loads.

The hardware necessary to accomplish the desired control strategies is readily available from commercial sources. All the proposed strategies are being practiced with a high degree of success at many trickling filter plants. Table 8 highlights the benefits of automating a trickling filter's recirculation rate, as well as some of its limitations and recommended uses.

Table 8. TRICKLING FILTERS

Control Method	Benefits and Potential Savings	Advantages	Disadvantages
Once through		Simple	Limited to low-rate filters
Constant recirculation	Protects filter	Simple; provides substrate during low flow	Higher equipment and energy costs
Throughput kept above minimum by controlled recirculation	Protects filter; saves power	Same as above	Higher equipment costs
Flow proportional recirculation above minimum flow	Provides maximum efficiency	Permits high rate of operation with predictable results	Requires variable-flow recirculation system, which in turn requires somewhat more maintenance

SECONDARY SLUDGE PUMPING

Introduction

Solids separation by gravity is the usual technique for removing secondary sludge in the effluent from the biological filters and the activated sludge aeration tanks. Generally, the comments made about primary clarifiers apply also to secondary clarifiers, but, in addition, the successful operation of secondary clarifiers must take the following into consideration:

- Sludge removal following the activated sludge process is limited by the return sludge requirements.
- It is especially important that secondary sludge be removed promptly and either returned to the process or processed for disposal. Secondary sludges are highly unstable and will become septic quickly, which has a deleterious effect on the activated sludge process.
- While sludges from biological filtration are small in volume (except during periodic filter unloading), they must be removed regularly in order to minimize potential septicity.

Control objectives, while similar to those for raw sludge, are dictated by the requirements of highly sensitive secondary treatment processes. Maintaining maximum working sludge density must be tempered by the ever-present specter of septicity.

Secondary sludges associated with the activated sludge process are subject to a condition known as "bulking," which will completely upset the entire secondary settling process. Since the sludge will not settle, none of the usual subsequent processes will function. This condition can be detected by a sludge level indicator, and appropriate measures should be taken to correct it. It is evident that control is essential:

- Because of the close relationship between the secondary sludge removal process itself and the activated sludge treatment processes
- In order to minimize the pumping of low solids secondary sludges
- Because the clarifier effluent must be void of settleable matter; otherwise, a high chlorine demand and poor effluent quality will result.

The final control element is a function of the secondary process. In the case of a biological filter, the usual element is a sludge pump. However, when the secondary process is activated sludge, the element can be a sludge pump, diverting valves, or a combination of both.

Control for secondary sludge removal in the case of biological filters is essentially the same as for raw sludge. The pump is started and stopped in response to the selected control parameter, and is usually powered by a fixed-speed motor.

On the other hand, the activated sludge process requires a more sophisticated final control element because the settled biomass must be returned in some regulated proportion. If the return sludge is not sufficiently concentrated, the MLVSS and the aeration tank residence time will decrease. Further attempts to build sludge age by increasing the recycle ratio will overload the secondary clarifiers, which will lead to additional process deterioration. Since the secondary process produces excess solids above return requirements, an appropriate amount of sludge also must be wasted to keep the process in equilibrium. The operating criteria for secondary clarifiers are as follows:

- Secondary sludge must be returned to the aeration process as required.
- Excess secondary sludge must be removed for treatment and disposal.
- Because secondary sludges from the activated sludge process deteriorate rapidly, secondary sludge must remain in the final clarifier as briefly as possible.

Control Strategies

Control of secondary sludge withdrawal from a biological filter system is relatively simple, since the process itself generates only small quantities of settleable solids in the normal day-to-day operation. Low-rate filters typically slough their solids seasonally, whereas high-rate filters continuously discharge small quantities of solids. Inasmuch as the biological sludge produced is not returned to the process but, instead, is directed to treatment and disposal, a slight anaerobiosis of the sludge is no problem. The usual control scheme consists of a timer to operate the sludge pump (similar to the time cycle control strategy used for primary sludge pumping).

While there are many variations of the control scheme cited above, no installations use the more sophisticated techniques, such as sludge waste interface detection or control based on sludge density. Apparently, both experience and engineering judgment indicate that such methods are unnecessary.

Successful operation of secondary clarifiers for the activated sludge process requires producing an adequate supply of return sludge and a high quality effluent that is essentially free from suspended solids. At times the dense sludge and high quality effluent goals may conflict; when this happens, the operator must decide which goal should be given the higher priority. Usually, producing an effluent free from settleable matter is considered more important than withdrawing a dense sludge. The F/M control section (described previously) details the specific designs and instrument loops associated with returning the proper amount of sludge. Because of the similar physics, equipment, and geometry used for primary and secondary clarifiers, the same control strategies and instrument diagrams apply to both types. Different processing objectives, however, change the desirability of specific control strategies.

Unlike primary clarifiers, most sludge withdrawal pumps for activated sludge processes are variable-speed devices so that they can return sludge to the aeration basin on a continuous basis. The most common technique consists of regulating sludge withdrawal, wasting, and recycling on the basis of the influent flow (as illustrated in Figure 18). The influent flow is measured and transmitted to a ratio controller that sets the index of the return sludge loop, consisting of a flow-measuring element, transmitter, controller, and final control element. Waste sludge is also regulated by a separate control loop that has the same equipment as the return sludge loop. Operating experience, laboratory data, or an F/M controller will determine the fraction of settled sludge returned.

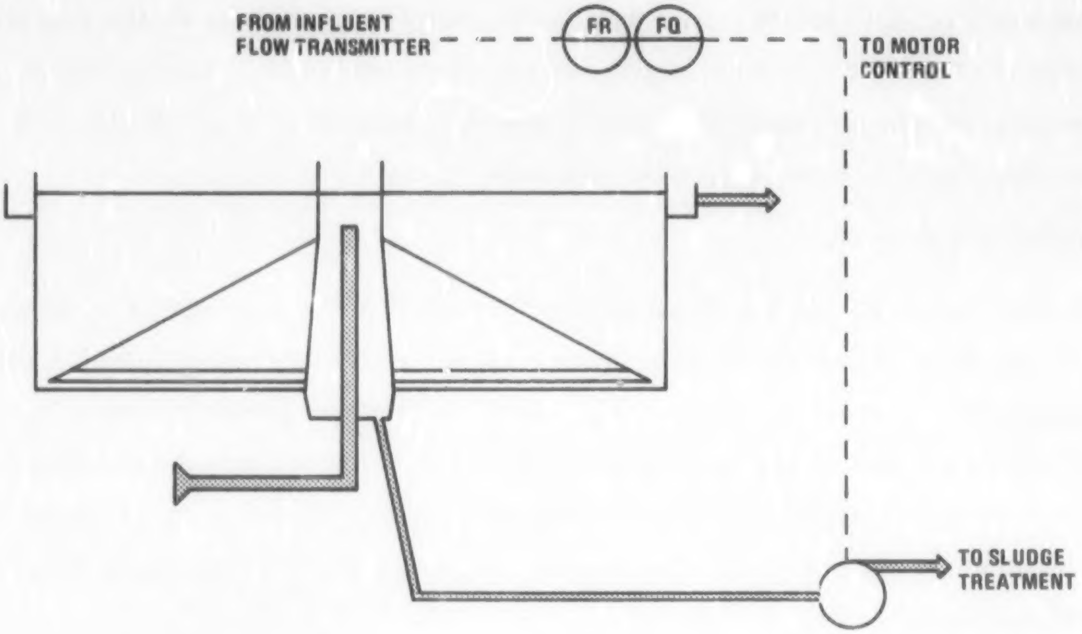


Figure 18. Flow proportional sludge pumping.

In a well-operated activated sludge system, the settling properties and volume of settled sludge change very slowly because the overflow rate of the clarifier is high (that is, the capacitance of the clarifier is large). Accordingly, the flow proportional system should provide satisfactory regulation. Flow proportion sludge recycle and sludge wasting control saves 30% of the labor required by manual operation, and effectively copes with flowrate perturbations.

One untried but potentially useful control strategy is to regulate the mass rate of returned sludge. If a sludge density meter monitors the recycle stream, then the product of sludge density and flow rate yields the mass rate at which microorganisms are being returned to the aeration basin. This sludge return loop can be cascaded with influent flow rates or incoming loads so as to maintain a proper F/M ratio.

Another possibility involves using sludge level control to regulate the wasting loop and either flow proportion or mass rate to modulate the return sludge. Although sludge interface devices are unproven, once this sensor demonstrates enough reliability, the strategy should be tested under field conditions.

Application Notes

Control system failures for secondary clarifiers result in the same types of problems as experienced in primary clarifier control failures. In the case of activated sludge, sludge pumping control failures strongly impact the treatment processes until ultimately the process itself could also fail. Continuous sludge withdrawal and return pump operation, which rarely occurs, will pull out the sludge blanket and eliminate the benefits of sludge recycle. Downstream sludge thickeners and stabilizing devices will soon become overloaded and fail.

Hardware availability and reliability are important factors, together with proper installation and preventive maintenance, in selecting potential control strategies. Suitable sludge flowmeters, sludge density monitors, and variable-speed sludge pumps are commercially available. Accordingly, flow ratio and mass rate control strategies can be readily implemented. Sludge interface detectors, although unproven in this type of service, should be shortly available and sufficiently reliable to use in an automatic control loop. Many municipal plants are successfully using sludge ratio, mass rate, and time cycle control strategies for automating their sludge return and wasting operations.

CHLORINE DISINFECTION

Introduction

Chlorine gas is widely used to disinfect industrial and municipal water and wastewater because it is effective, easy to apply and inexpensive. It has a high toxicity for microorganisms responsible for waterborne diseases. In certain applications (such as storm-water treatment and remote package treatment systems), hypochlorite solutions or bleaching power provides a safer, more flexible means of disinfection than gaseous chlorine. Hypochlorite and bleaching powders, however, are more expensive and deteriorate with time. Other disinfection agents such as ozone and ultraviolet irradiation are used only occasionally because they are in general more expensive.

Chlorine gas, which is very soluble in water (0.7% at 20°C and 1 atm), hydrolizes rapidly to form hypochlorous acid when dissolved in water:

$$Cl_2 + H_2O \rightleftharpoons HOCl + H^+ + Cl^-$$

At chlorine concentrations of less than 0.1% and pH values greater than three, the hydrolysis goes virtually to completion. The rate of kill has been empirically correlated to the 1.3 power of the residual chlorine concentration.

Although a rigorous accounting of all the forms and reactions of chlorine and compounds present in wastewaters is beyond the intended scope of this report, it is instructive to examine some of the principal reactions among chlorine, reducing agents, and ammonia. When chlorine is added to water-containing reducing agents such as hydrogen sulfide and nitrites, the hypochlorous acid reacts with the reducing agents to form chlorides, and no useful disinfection results. After satisfying this demand, further chlorine additions will result in the formation of chloramines. Chlorine stored in the form of chloramines is available for disinfecting purposes (this is usually referred to as combined available chlorine). Continued chlorination initiates the complete oxidation of the chloramines to nitrogen, nitrate and nitrogen trichloride (breakpoint chlorination). As a practical guide, because of the high ammonia concentrations present in wastewater, only combined residual chlorine is used as a disinfectant.

Chlorine residual of 0.5 mg/l after a 15-minute contact period is generally necessary to ensure the adequate disinfection of sewage treatment effluent. The wastewater must be thoroughly mixed with the chlorine and then processed in a properly designed contact chamber. Because of the variable sewage flowrates and reducing agent concentration, the chlorine feed rate must be adjusted continually in order to maintain a constant residual chlorine level.

A good automatic chlorine control system should supply sufficient chlorine to ensure adequate disinfection and should also minimize chlorine consumption. Furthermore, the control system should be safe, easy to maintain, and compatible with the environment of a wastewater-treatment plant. Fortunately, these broad control objectives can be accomplished with several widely used and well-proven chlorination control strategies, which range from simple flow proportional to the complex compound plus postcontact residual chlorine control. Chlorine feed rate is always the manipulated variable, and the final control element may be a valve, a metering pump, or a chlorine feed loop. The selected control strategy then sends a chlorine feed rate signal to the final control element, which implements the final action.

Flow Proportional Control

The most widely used control strategy regulates chlorine feed in proportion to plant flow, as depicted in Figure 19. To make this control strategy effective, the flowrate should be measured directly at the head of the mixing chamber or as closely as practical to the point of chlorine application. A common error has been to use the flow measured at the plant's headworks rather than the flow at the chlorination site. Although the long-term average flow will be the same, the immediate difference in flowrate can be large. For example, returning back flush water or digester supernatant will momentarily increase the down-stream flowrates. When properly designed, the flow-proportioned control system will automatically adjust the chlorine feed rate in accordance with the flow variation so as to maintain a constant ratio between the chlorine added and the wastewater flow. The plant operator should take samples periodically of the chlorinated effluent for bacterial and residual chlorine analysis, and, on the basis of this information, he should manually adjust the ratio of chlorine added to wastewater flow.

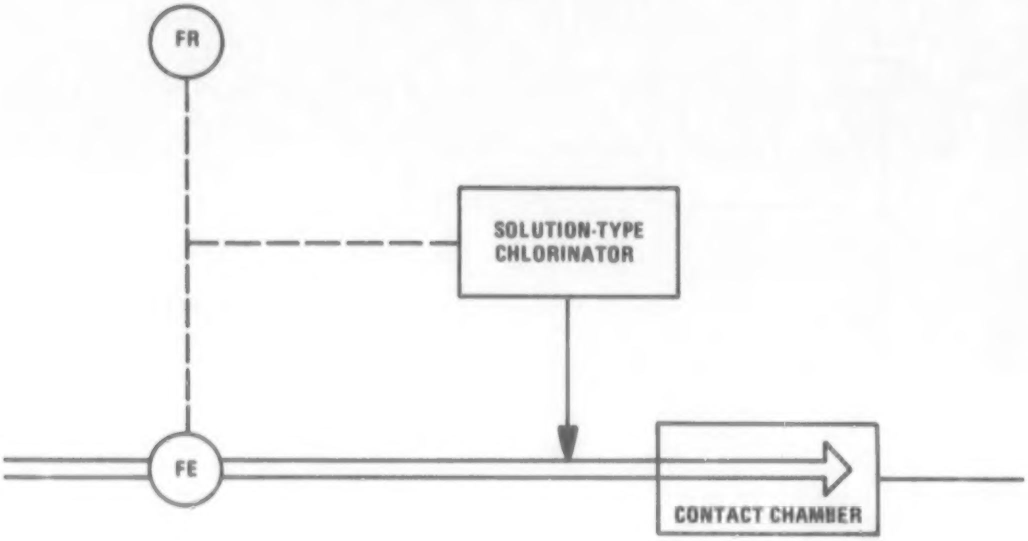


Figure 19. Simple flow-pacing chlorination control loop.

Flow-proportioned chlorination control is reliable and simple to implement and maintain; it also responds rapidly to flowrate perturbations. Since this strategy neglects changing chlorine demands on an automatic basis, flow-proportioned chlorination is not well suited to applications where the chlorine demand changes frequently or where subsequent dechlorination processes are employed. Nevertheless, flow proportion control is adequate for the many small plants that use small-to-moderate amounts of chlorine.

Compound Chlorine Control

With the availability of reliable residual chlorine analyzers, the ratio between the chlorine feed rate and wastewater flowrate can be adjusted automatically on the basis of on-line residual chlorine measurements (Figure 20). Since the chlorine dose depends upon the quantity of reducing agents present, the ammonia concentration, the organic load, and the amount of suspended solids, the chlorine demand changes significantly in the course of a day in many wastewater-treatment plants. Accordingly, the amount of chlorine per volume of wastewater necessary to achieve a given residual

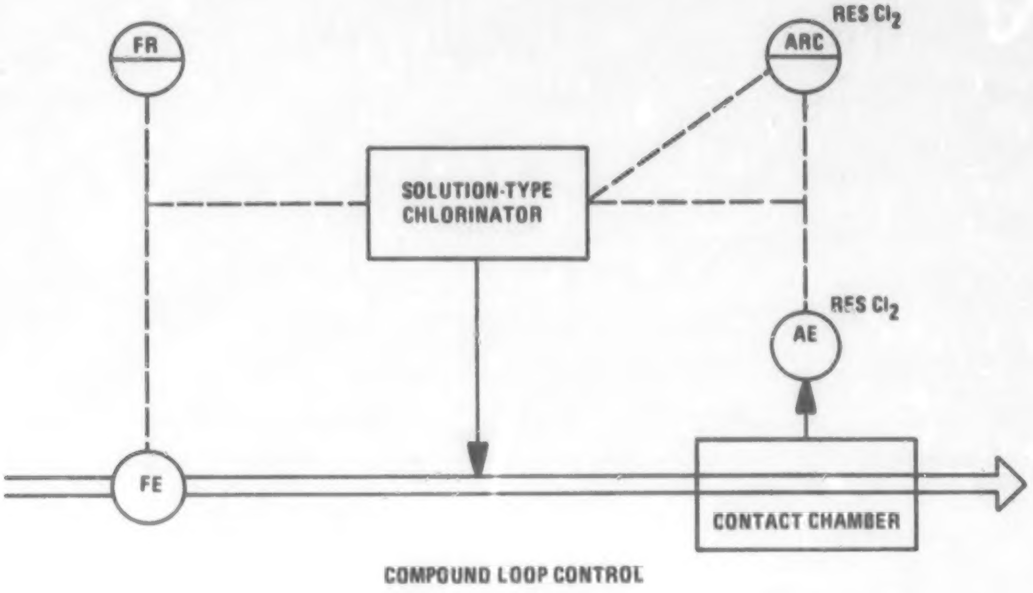


Figure 20. Typical compound chlorination control loop.

chlorine level also changes significantly in the course of a day. With the onset of a high chlorine demand period, the residual chlorine drops below the desired level. The residual chlorine analyzer then sends the concentration data to the controller, where an increased chlorine flowrate command is generated and acted on by the final control element. This feedback system automatically increases or decreases the flow ratio until the desired residual chlorine level is obtained.

Compound chlorination control accurately modulates the chlorine application and so maintains the desired residual chlorine level. Many wasterwater-treatment facilities are successfully using this control strategy today. Although the automatic control equipment is more expensive than simple flow-proportioned control equipment and requires more maintenance, reliable equipment is commercially available, with its higher costs offset by the chlorine gas savings obtainable by tighter control. Consequently, compound chlorination control is considered suitable for medium-to-large plants that use significant amounts of chlorine and/or experience time-varying chlorine demands, and for plants that use dechlorination processes.

Double Compound Control

Compound control, however, does pose some difficult problems because most standards or codes require that a prescribed residual chlorine be maintained after at least a 15-minute contact time. Accordingly, feedback residual chlorine control systems that have potential 15-minute lags are prone to instabilities. If a feedback control system is to perform adequately, the loop time must be within a 3-to-5 minute range; this means that the residual chlorine must be determined shortly after mixing. The difficulty of relating the control residual to the residual at the end of the proper contact period is best handled by a second residual chlorine analyzer that records the residual after contact. To ensure an adequate residual, it may be necessary to use a postcontact residual chlorine analyzer that readjusts the chlorine application rate at the head of the contact chamber as shown in Figure 21.

Both the compound and double compound control strategies use pacing or elementary feedforward as the primary action, and residual chlorine as a secondary trimmer. Since pacing is a highly stable control mode and the chlorine demand normally changes slowly, compound control loops are very stable.

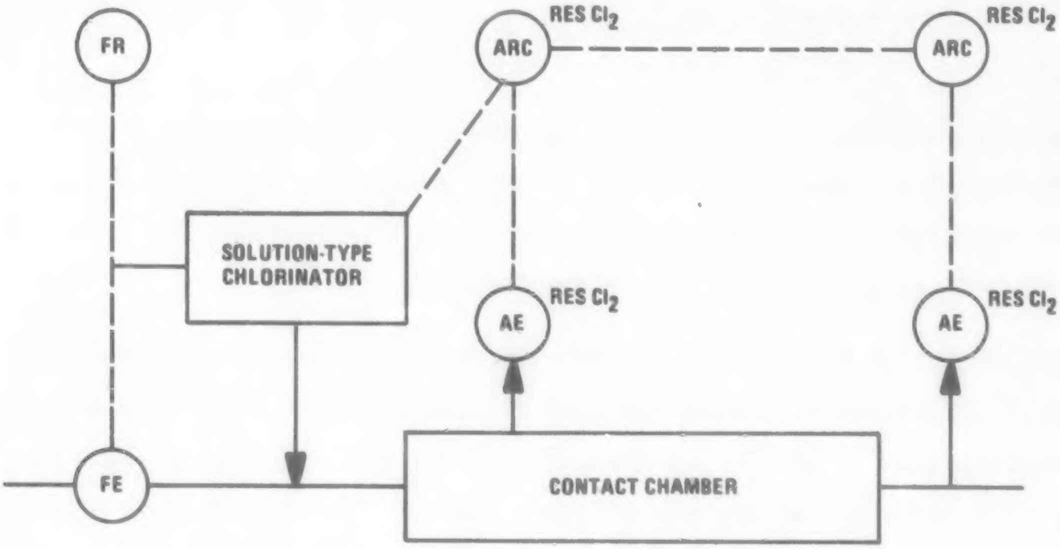


Figure 21. Double compound loop Cl_2 control.

Application Notes

The control dynamics of the chlorination process are dominated by both reaction time and large delays. Control will be optimum with proper design to ensure that: 1) the chlorine water stream is well mixed into the main flow, 2) the flow signal is properly represented at the point of application, and 3) when an analyzer is used, the measurement lag does not vary appreciably with the plant's throughput rate.

The success of the compound controllers depends on accurate residual chlorine measurements; these analyzers require periodic maintenance and calibration. Accumulated user experiences indicate that about 140 manhours a year must be allocated for servicing each residual chlorine analyzer.

Chlorination control system failures can have a nearly catastrophic effect on the operation and safety of a wastewater plant. A failure that results in the cessation of chlorine feed permits unchlorinated effluent to be discharged into the receiving water. Other failures can result in overchlorination (residual chlorine analyzer failures are principally responsible for this type of failure). Chlorine will be fed into the wastewater up to the capacity of the chlorination equipment, regardless of demand or residual. The impact of gross overchlorination depends upon the nature and size of the receiving water.

There are hundreds of flow proportional chlorination systems in service today. A recent survey found that 94% of the users were satisfied with the system's performance. Flow-proportioned control is particularly well suited to small plants, where it is usually difficult to obtain skilled maintenance service.

About one-third of the existing wastewater-treatment plants use compound control to regulate the gaseous chlorine feed rate, whereas only a handful of plants use double compound control. Typically, the newer and larger plants are more apt to use compound control because of their need for tighter residual chlorine control. A survey of some 13 plants that practice compound control disclosed that 77% of them were satisfied with the compound control loop's performance.

The automatic residual chlorine control devices that are presently available are field proven and ensure the proper chlorination of wastewaters, especially after secondary treatment. Occasionally, chlorination control of raw sewage, storm water, or combined sewage may fail because the residual analyzer becomes plugged with debris. Compound control systems can pay for themselves in chlorine saving, while simultaneously ensuring a facility's compliance with discharge standards.

The data presented in Table 9 was obtained from the survey as reported by Molvar et al. (1) and highlights the benefits and limitations of the chlorination control strategies.

Table 9. DISINFECTION VIA CHLORINE ADDITION

Control Method	Benefits and Potential Savings (1)	Advantages	Disadvantages
Fixed rate		Simple and reliable	Can overdose or underdose as flow changes
Flow proportion	15% labor saving and 25% chlorine saving over basic method	Reliable and well-established	Requires some maintenance
Residual chlorine feedback	15% labor saving and 50% chlorine saving over basic method	Well-established and produces good effluent residual control	Requires some maintenance but more instrumentation; poor performance when flow changes rapidly
Compound control	15% labor saving and 50% chlorine saving over basic method	Well-established and produces excellent residual control	Requires some maintenance and instrumentation
Double compound control	15% labor saving and 55% chlorine saving over basic method	Best available control of final residual; especially useful whenever nature and strength of influent vary widely	Requires considerable maintenance and instrumentation

CONTROL OF ANAEROBIC DIGESTERS

Introduction

Anaerobic digestion is a complex biological process which converts organic matter to methane and carbon dioxide in the absence of molecular oxygen. This process is widely used in the stabilization of domestic and industrial wastewater sludges. The goal of anaerobic digestion, as well as any other sludge stabilization process, is to produce an easily dewaterable sludge which can be safely disposed of without environmental nuisances or hazards. Anaerobic digesters, however, have a reputation of being unstable, unreliable, and troublesome. This belief is primarily due to improper operation and control, rather than any inherent instabilities of the anaerobic digestion process.

Although relatively little is known about the metabolic processes that occur during anaerobic digestion, certain operational concepts and practices have been established and accepted. For example, such factors as volatile acids concentration, alkalinity, pH, retention time, biomass concentration, loading rates, and temperature strongly influence the stability and operational efficiency of anaerobic digesters. Process control technology offers a method of improving the reliability of anaerobic digesters by automatically regulating some of the above-mentioned factors. The automatic monitoring associated with process control also would measure process conditions and tend to eliminate human errors. The following sections develop control strategies for the three most important process control parameters: temperature, pH, and methane gas production.

Temperature

Since the reactions taking place in anaerobic digesters result from the metabolic activity of heterogenous bacterial populations, the temperature effect on process efficiency is determined by the response of the bacterial species present. Temperature ranges for the optimal growth of microorganisms can be divided into three regions:

psychrophilic (<20° C), mesophilic (20 to 45° C), and thermophilic (>45° C) (9). Although microorganisms are active from -5° C to about 80° C, specific microorganisms often have narrow temperature ranges in which they will reproduce. In fact, bacteria are classified as psychrophilic, mesophilic, or thermophilic bacterium according to the optimal growth temperature. A common taxonomic criteria also uses growth temperature ranges for a specific bacterium. Thus, temperature strongly influences the ultimate population that will actively grow in the digesters.

Most authorities agree that economical operation of anaerobic digesters occurs in the mesophilic and thermophilic zones, although much controversy exists on the pros and cons of mesophilic vs thermophilic digestion (10, 11).

The rate of a chemical reaction doubles for every 10° C increase in temperature. Over a narrow temperature range this is approximately true for biological reactions. Also, it has been demonstrated that sudden temperature changes are detrimental to anaerobic digesters (12). Since it is important to maintain the digester at a temperature which allows stabilization to proceed at the highest possible rate, good temperature control is essential.

Older digesters have heating coils embedded in the digestion tank through which hot water is circulated to transfer energy to the digester liquor. Present designs pump the sludge to an external heat exchanger where the digester liquor is recirculated, resulting in higher thermal transfer efficiencies and in better mixing.

Since digesters are designed and operated as large-capacity processes with residence times from 10 to 30 days, simple two-mode feedback temperature control, as illustrated in Figure 22, keeps the digester's temperature within acceptable limits. The resistance or gas-filled temperature element, which must be located in a well-mixed representative region, measures the digester's temperature. When the temperature decreases below the lower setpoint (differential gap), the controller turns on the sludge recirculating pump. Accordingly, the digester's contents are heated and the temperature rises. When the temperature increases to the upper limit, the controller shuts

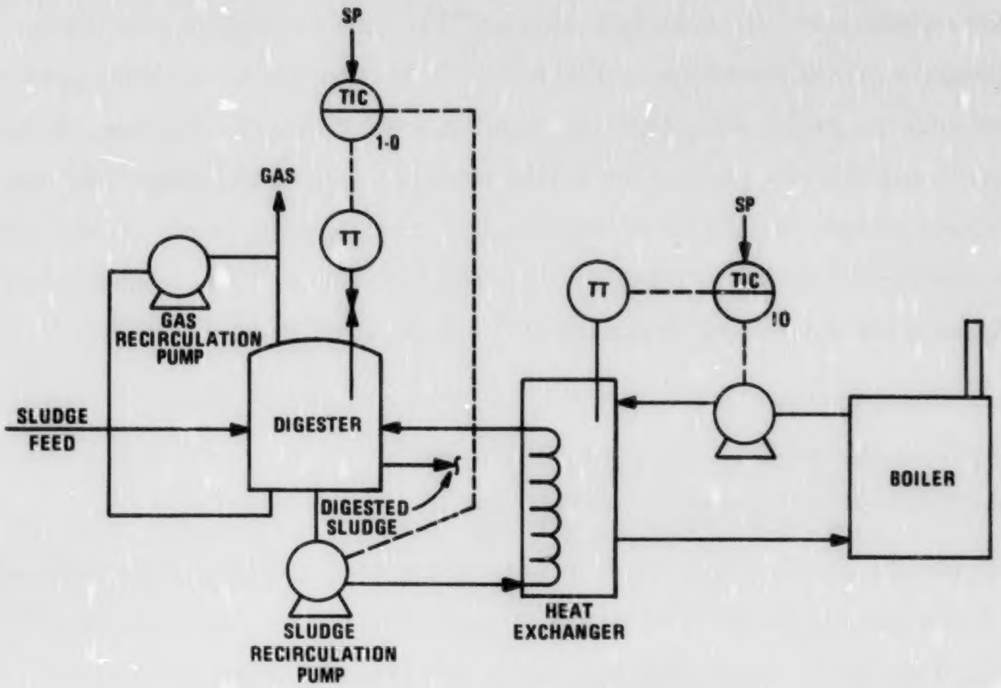

SP
TIC
1-0
GAS
TT
SP
TT
TIC
I0
GAS
RECIRCULATION
PUMP
SLUDGE
FEED
DIGESTER
BOILER
DIGESTED
SLUDGE
SLUDGE
RECIRCULATION
PUMP
HEAT
EXCHANGER

Figure 22. Digester temperature control.

off the recirculation pump. A second on/off thermostat controls the boiler or heater so as to maintain the proper hot-water temperature. Most commercially available temperature control systems can satisfy the anaerobic digester's requirement of maintaining a uniform temperature.

pH

pH is an important indicator of the condition of anaerobic digesters. Methane production results from two major groups of microorganisms: the acid-forming group, which is responsible for hydrolyzing the complex organics to simpler compounds (typically volatile acids), and a second group, the methane formers, which are sensitive to pH. Most reports indicate that the optimum pH is in the range of 6.8 to 7.2. If the pH is out of the 6.8 to 7.6 range, the methane bacteria become inhibited and sludge stabilization ceases.

The three common causes of digester failures (hydraulic, organic, and toxic overloadings) result in pH changes. The magnitude of the pH change depends upon the size of the overload and the alkalinity of the digester liquor. Hydraulic and organic overloadings produce a rapid decrease in pH due to the rapid buildup of volatile acids. Toxic materials preferentially kill the sensitive methane-forming bacteria which results in a gradual reduction in pH.

Since the methane bacteria become inhibited outside the 6.8 to 7.2 range, automatic pH control can eliminate the deleterious effects of pH upsets caused by hydraulic and organic overloadings. In a healthy digester the formation of bicarbonate alkalinity counterbalances the formation of volatile acids, and the pH remains constant. Because it is difficult to assess the cause of pH disturbances, a feedback pH control system (as shown in Figure 23) should hold the pH in an acceptable range. Usually lime is used for pH control of digesters. However, unless high intensity mixers are used proper dissolution and distribution of lime in digesters does not take place. It is not

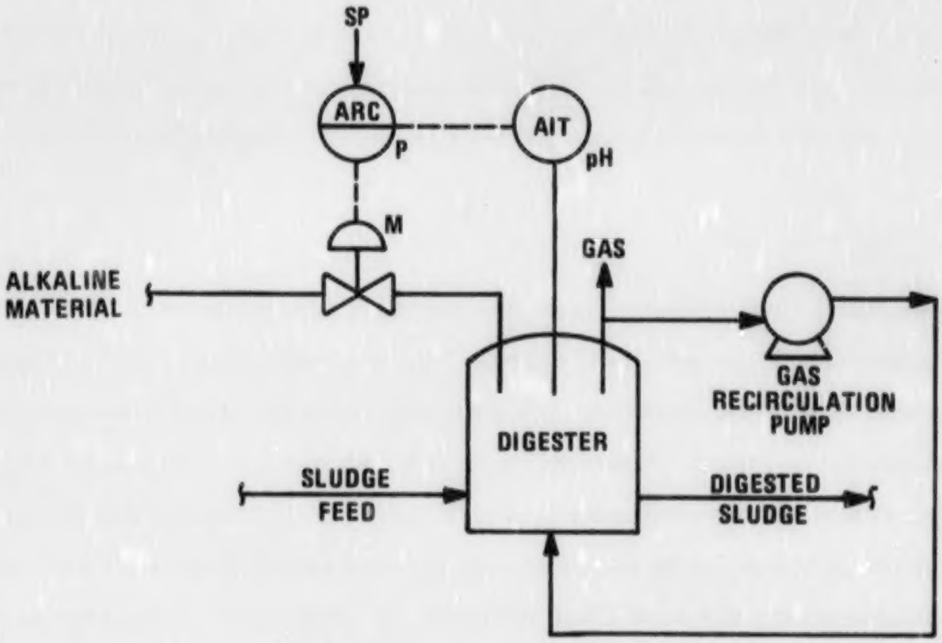


Figure 23. Digester pH control.

good practice to use NaOH because it rapidly reacts with CO_2 and could produce a partial vacuum. Sodium bicarbonate or sodium carbonate are the chemicals of choice. A recent stimulation study (13) shows that simple proportional control can keep the pH within acceptable limits.

The control strategy depicted in Figure 23 shows a pH feedback control loop for a well-mixed anaerobic digester. Since low pH's are responsible for most digester problems, provisions have been made only for the addition of alkaline reagents. When the pH drops below the set-point, the alkaline material is added at a rate proportional to the pH error. The addition of sodium hydroxide raises the pH to the acceptable range and thus provides a more favorable environment for stable digester operation.

The fouling nature of sludge liquor means the pH probes will become coated in a short time. Consequently, it will be difficult to keep an automatic pH control system operating correctly. The addition of in situ ultrasonic probe cleaning and installing the probe in a well-mixed region of the digester may eliminate or at least minimize the pH probe fouling problem. Each proportional pH control system should require about 60 man hours of maintenance per year for cleaning and calibrating the pH probes and tuning the proportional controller. Based on engineering estimates, automatic pH control of anaerobic digesters may allow a 10% increase in sludge processing.

Methane Gas Production

The goal of stabilizing sludge by anaerobic digestion can be achieved in only one way—the production of methane gas. The degree of organic removal is in direct proportion to the amount of methane produced. When the gas production trend is downward, the digestion process is failing. However, several events which are not related to process failures may cause variations in gas production. For example, decreasing the amount of sludge fed to the digester will clearly lower gas production. Also, temperature variations of only 2°F or 3°F will decrease gas production. In the forthcoming control development, it will be assumed that adequate temperature control exists so that decreasing methane gas production indicates instability and impending digester failure.

Since anaerobic digestion is a two-step process, a decrease in methane gas production implies that the number or reaction rate of the methane-forming bacteria is inadequate to convert the organic intermediates to methane and carbon dioxide. More methane formers must be added to the anaerobic digesters or the reaction rate must increase. In single-stage digesters, the methane population will increase by auto-catalytic growth as a result of excess available food. Because methane formers are slow growing, the digester cannot be fed; otherwise the process becomes inhibited by pH depressions. Increasing the digester's temperature also increases the growth rate and the rate of volatile acid destruction.

For single-stage digesters, accordingly, the rate of methane gas production can be used to automatically modulate feeding rates, temperature, and pH, as shown in Figure 24. If methane gas production falls and the rate of loading is unchanged and no toxic materials are present, then it would be desirable to increase the temperature. When the methane gas production drops, the measuring elements transmit this information to the control system, increasing the digester's temperature inversely proportional to the gas production rate. If this strategy does not correct the situation, digester feeding is terminated when the gas rate decreases below a minimum value. If the temperature control works, the methane gas production rates rise to their usual values, the temperature is reduced gradually to its normal levels and the digesters are once again fed at a uniform rate. Unfortunately, the methane formers grow slowly and several days are necessary to restore a single-stage digester.

With two-stage digesters, the most effective control strategy involves recycling settled sludge (methane bacteria) from the second to the first digester. This strategy leads to a buildup of methane-forming bacteria when gas production drops. The methane production data by means of a feedback controller regulates the sludge

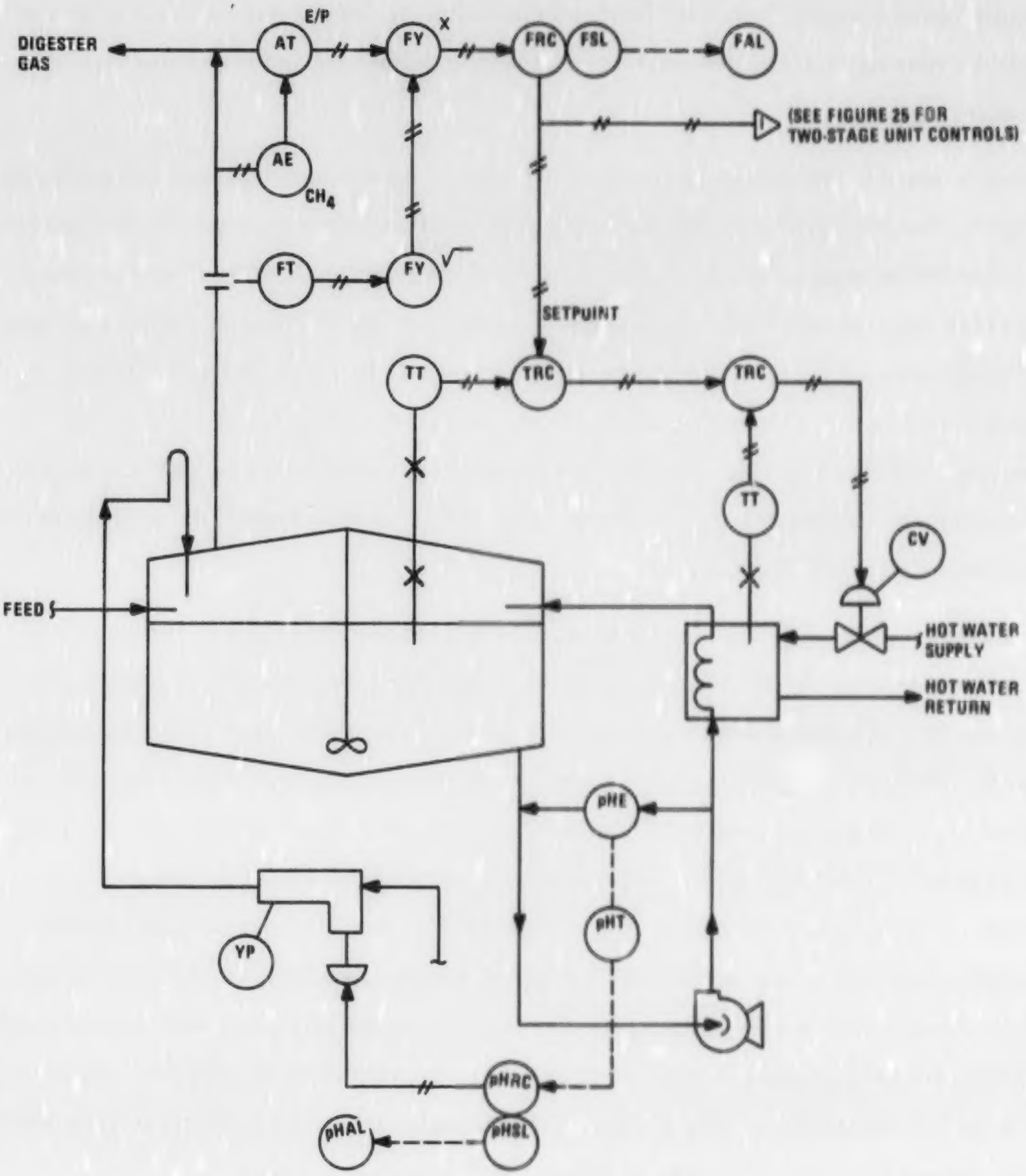

DIGESTER GAS
AT
E/P
FY
X
FRC
FSL
FAL
(SEE FIGURE 25 FOR TWO-STAGE UNIT CONTROLS)
AE
CH4
FT
FY
SETPOINT
TT
TRC
TRC
TT
CV
FEED
HOT WATER SUPPLY
HOT WATER RETURN
pHE
pHT
YP
pHRC
pHSL
pHAL

Figure 24. Combined pH, temperature, and digester gas control for single-stage unit.

recycling, as illustrated in Figure 25. Since the methane production is a linear function of the methane bacteria concentration and the control objective is to keep the methane production rate above some minimum value, a simple on/off controller should provide adequate corrective control action. When the methane gas production drops, the sludge recycle pump is turned on until the methane rate rises above the prescribed value. The recycle sludge pump is then turned off. To be successful, the second-stage digester must contain a sufficient supply of methane bacteria; otherwise, increasing the sludge recycle rate will simply decrease the hydraulic retention time and performance will deteriorate. In most cases it is desirable to cascade this control scheme with temperature and pH controllers, as was done in the single-stage digester.

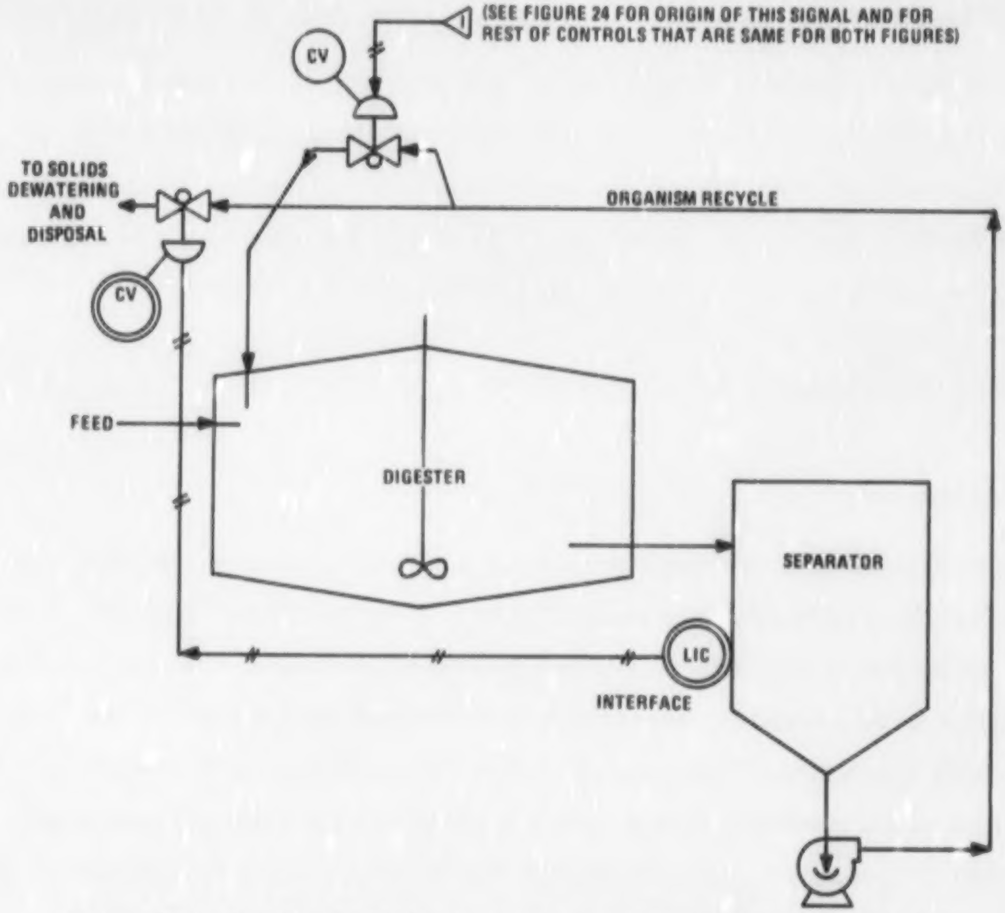

(SEE FIGURE 24 FOR ORIGIN OF THIS SIGNAL AND FOR REST OF CONTROLS THAT ARE SAME FOR BOTH FIGURES)
CV
TO SOLIDS DEWATERING AND DISPOSAL
ORGANISM RECYCLE
CV
FEED
DIGESTER
SEPARATOR
LIC
INTERFACE

Figure 25. Combined pH, temperature, and digester gas control for two-stage unit.

Both these methane gas production control strategies have never been tried in a wastewater-treatment plant. Reliable methane or carbon dioxide analyzers and gas flowmeters are readily available and have successfully demonstrated their suitability for wastewater-treatment projects. Undoubtedly many plants are using methane gas production rates in formulating their operating practices, but these control strategies are being practiced on a manual rather than automatic basis. Simulation studies have shown that methane bacteria recycling based on methane gas production rates will prevent most digester failures automatically. In the absence of specific performance data, the proposed control strategy for single-stage digesters should permit processing about 15% more waste sludge than a manually controlled digester. For two-stage digesters, the proposed control action should increase sludge processing by about 25%.

Since sludge stabilization accounts for approximately 40% of a treatment plant's capital cost, anaerobic digesters, the most commonly used sludge stabilization process, should be carefully considered during plant automation and process control. The proposed control strategies summarized in Table 10 have the potential for saving capital and operating expenses, and for increasing the digester's reliability.

SLUDGE CONDITIONING FOR VACUUM FILTRATION

Introduction

Chemical conditioners are required to effectively dewater activated sludge and primary and digested sludges by vacuum filtration. Polyvalent metal ions such as Al (III) and Fe (III) or synthetic organic polyelectrolytes (cationic, nonionic, or anionic) are added to the sludges in order to structure them properly for dewatering. These chemicals attach themselves to the discrete sludge particles and form a bridge to other individual particles. The conditioned sludge now has sufficient structural strength and porosity to allow the rapid escape of water under a vacuum-driving force.

A recent laboratory study (14) showed that the supernatant pH and the sludge solids concentration significantly affect the conditioner dosage required to enhance sludge

Table 10. ANAEROBIC DIGESTER CONTROL STRATEGIES

Description	Potential Savings	Confidence	Advantages	Disadvantages	Remarks
Temperature	10% more throughput	High	Maintains temperature within 1°C of desired value. Presently available equipment is satisfactory.		Temperature control should be used on all digesters.
pH	15% more throughput	Low	Keeps pH within the optimum range of 6.8 to 7.6. Compensates for organic overloading.	pH probe fouls easily due to nature of digester sludge. In fact, this strategy has never been tried in a municipal plant.	pH probes must be located in a well-mixed region. Ultrasonic cleaning may be helpful. Untried.
Methane gas production (single stage)	15% more throughput	Moderate	Acts as an early warning system with automatic corrective action for toxic overloading.	Slow recovery time (digester is out of service for this time). Severe corrosion problems with gas metering.	This control strategy is useful when toxic overloading occurs frequently. Should be cascaded with temperature control.
Methane gas production (two stage)	25% more throughput	Moderate	Automatically compensates for toxic overloading by introducing new methane-forming cells.	Must ensure an adequate supply of concentrated sludge for return to first digester. Severe corrosion problems with gas metering.	This control strategy is useful when toxic overloading occurs frequently.

dewatering by vacuum filtration. Since process disturbances can change the pH and the mass loading of settled sludge, automatic control systems that adjust chemical feed rates so as to keep the sludge cake production at a maximum level and to produce a cake with a minimum moisture content (60% to 70%) are examined.

pH Control Systems

The hydrogen ion concentration will affect the surface charge of the sludge particle as well as the properties of the conditioners. The optimum pH for effective conditioning of sludge is 6.0 to 7.0 for Fe (III) and 4.5 to 5.5 for Al (III). Although a feedback control system that acts on the conditioned sludge's pH would provide ideal control (refer to previous discussion), pH is very difficult to measure continuously in thickened waste sludge. Accordingly several alternative pH control systems have been proposed.

Rather than measure the pH of the conditioning sludge, the automatic pH control system responds to changes in the filtrate's pH, as depicted in Figure 26. Since the solids do not affect the pH, measurements of the filtrate's pH are indeed a measure

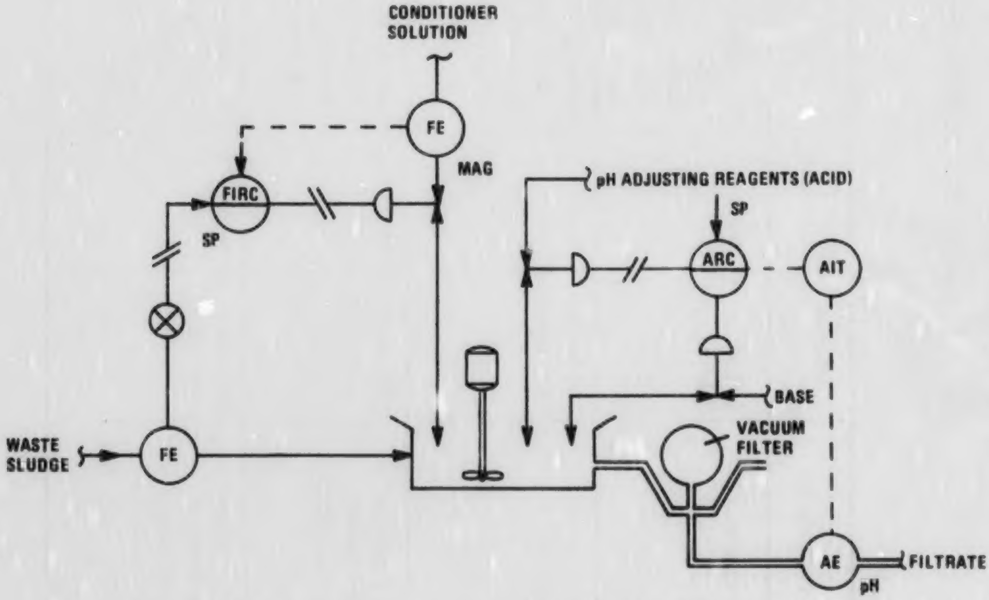


Figure 26. Feedback pH control based on filtrate pH.

of the slurry's pH. However, the system has a deadtime of several minutes, which may cause some control difficulties. Fortunately, the alkalinity of the sludge slurry and chemical demand changes so slowly that most of these systems work reasonably well.

A second method of pH control regulates the pH of the conditioning chemical solutions, as shown in Figure 27. The success of this method pivots on the constant alkalinity of the sludge slurry and the ability of the flow proportional dosing system to perform accurately.

Although none of the plants visited in a recent nationwide survey practiced any form of automatic sludge-conditioning pH control, it is estimated that automatic pH control would increase a vacuum filter's productivity by about 20% over a well-run manual operation.

Dosage Control

The data reported by Tenney (14) clearly shows that a stoichiometric relationship exists between the sludge mass and the dosage of chemical conditioners. Moreover,

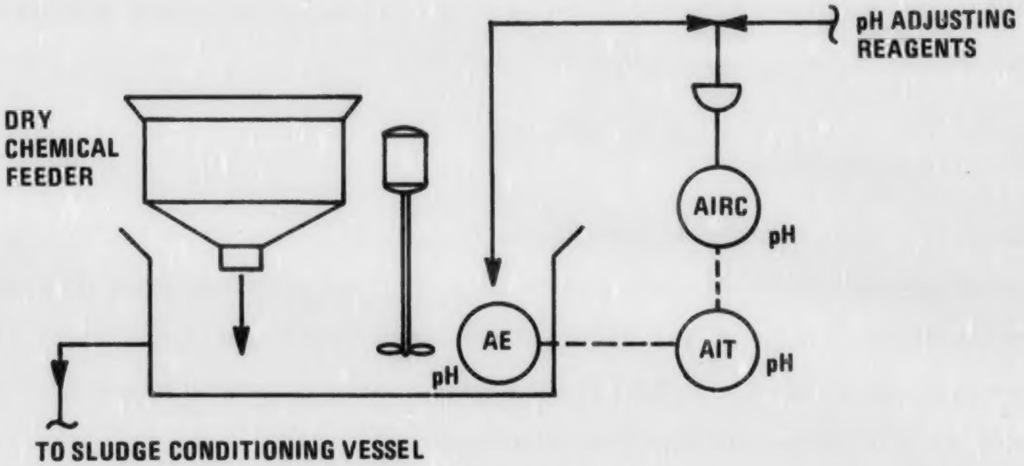


Figure 27. pH control of conditioning chemical solutions.

this relationship appears to be linear, as shown in Figure 28. If the sludge concentration remains essentially constant, then the conditioning chemicals can be added on a flow proportional basis, as illustrated in Figure 29. However, where the solids concentration also changes, the conditioners should be added on a mass proportional basis. This strategy would require a density-measuring device such as a nuclear or ultrasonic sludge meter, as shown in Figure 30, which also depicts the other elements of this control system. Both the flow proportional and mass proportional control systems require accurate flow rate-monitoring devices. Unfortunately it is difficult to measure flow rates reliably at the low velocities usually used in transporting thickened sludge. Specially designed flowmeters, which became available only recently (such as BIF Solid Bearing Fluids Meters), should be used.

Automatic dosage control is rarely used in wastewater-treatment plants today; instead, constant rate chemical feeding is used most frequently. Constant dosage control will probably reduce chemical consumption by about 15% and increase throughout by 5 to 10%. Mass proportional dosage control should reduce chemical consumption by 20 to 25% and increase the filtering capacity by about 10%.

Before extensive efforts are undertaken to demonstrate these potentially useful but untried control strategies (which also use new and relatively untried flow-measuring elements), the extent of need must be carefully appraised, since today vacuum filtration is only one of many acceptable techniques for disposing of waste sludge.

SLUDGE DEWATERING

Because vacuum filtration reduces the water content of sludge from 95% to 20%, it is widely used to reduce the water content of sludge prior to its final disposal by means of landfilling or incineration. In today's wastewater-treatment plants, vacuum filters are operated intermittently for several days/per cycle to dewater a backlog of waste sludge. The fundamental filtration equation is:

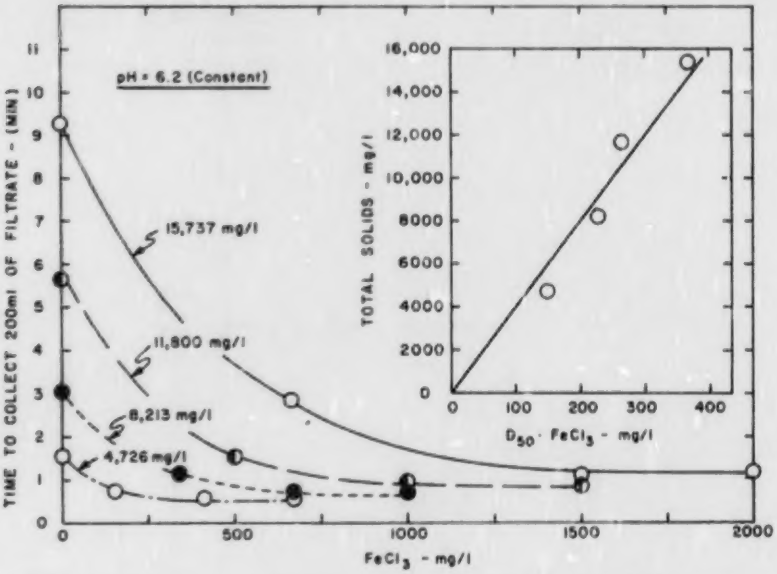


Figure 28. Effect of increasing sludge solids concentration (dry basis) on requisite conditioner dose. Increasing solids concentration (or surface area) increases requisite conditioner dose proportionately. A definite stoichiometry exists between sludge solids concentration and requisite conditioner dose, as shown by insert.

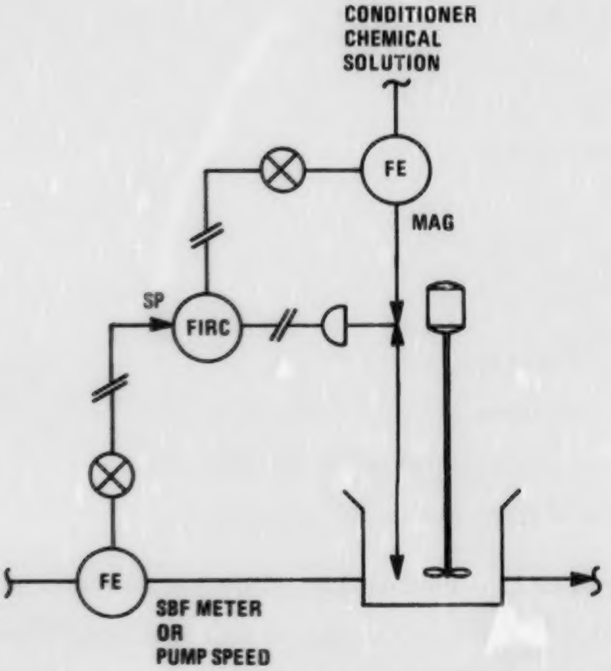

CONDITIONER
CHEMICAL
SOLUTION
FE
MAG
SP
FIRC
FE
SBF METER
OR
PUMP SPEED

Figure 29. Flow proportional chemical feed system.

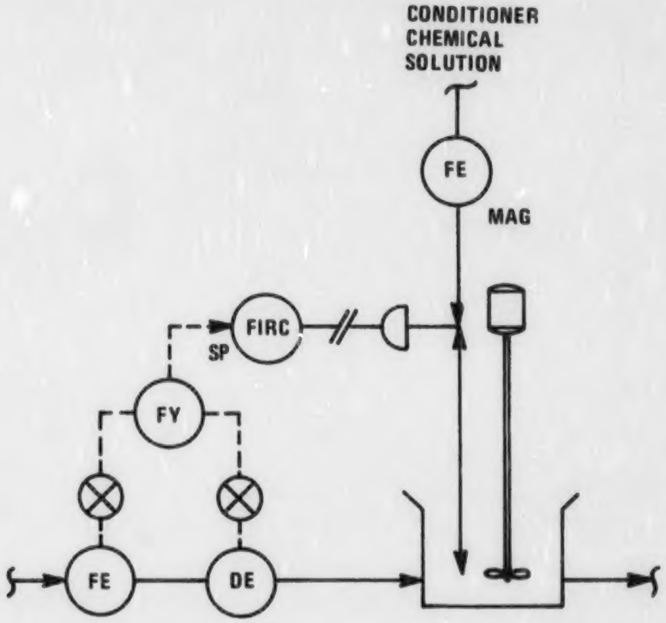

CONDITIONER
CHEMICAL
SOLUTION
FE
MAG
FIRC
SP
FY
FE
DE

Figure 30. Mass loading chemical feed system.

$$\frac{dV}{dt} = \frac{PA^2}{\mu(rcV + RmA)} \qquad (22)$$

where:

V = volume of filtrate

t = time

P = pressure

A = area

μ = viscosity of filtrate

r = specific resistance

c = weight of solids/unit volume of filtrate

Rm = resistance of filter media

This equation shows that the performance of the vacuum filter as measured by the filtration flow rate $\left(\frac{dV}{dt}\right)$ depends on the stream variables of specific resistance (r) and solids concentration. Proper chemical conditions by means of several automatic control systems should keep the resistance at a stable low value. Filter performance also depends on the machine variables of pressure drop, renewed surface area rate (rotational speed), and sludge feed rate.

From a control standpoint, the objectives of sludge dewatering are to maximize the production of dewatered sludge and to minimize the blinding of filter media in spite of any process disturbances. The latter objective is largely a function of the filter media and is not a major problem in view of contemporary designs and materials. Process control is necessary as a means of establishing sludge flow and the filter's rotational speed. Usually the supply trough is maintained at a constant level by pumping the sludge into the supply trough and subsequently returning the excess through a gravity overflow.

In small and medium-size plants, the vacuum filter may operate only 2 or 3 days/week. Sludge feed rate and rotational speed control eliminates the need for constant operator attendance, and also results in a more uniform sludge cake. Unfortunately the limitations involved in measuring low-velocity sludge flows entail a serious impediment to success. Because the present control technique (of maintaining the sludge trough level by pumping an excess and then returning it through an overflow) is both simple and effective, possible improvements of this technique are not likely in the near future.

There is an alternate technique: starting and stopping the sludge feed pump in order to maintain an acceptable range of levels in the supply trough, but this method triggers a potential problem, the possibility that the feed line will become plugged during the feed pump's off-time.

Hardware for the reliable level measurement, flow measurement, and final control elements is only partially available. Unfortunately any limitation of the sludge flow by either variable-speed pumps or valves promotes plugging of the feed line. A computer offers no advantages to this process, and neither are complex formulations or elaborate equipment sequencing required.

The failure of vacuum filter control systems in present-day service results in flooding the supply trough and/or starving the filter, thus taking the vacuum filter out of service until the situation can be corrected. Normally no permanent damage results, however.

The current technique of pumping excess sludge is essentially self-regulating, and the advisability of future improvements remains questionable because the added equipment would require extensive maintenance and introduces operating problems of its own.

Table 11 highlights the advantages, savings, and limitations of vacuum filtration control schemes.

Table 11. SLUDGE DEWATERING—VACUUM FILTER

Control Method	Benefits and Potential Savings	Advantages	Disadvantages
Supply trough level control (overflow)	Uniform sludge cake production	Simple and effective	Excessive pumping; complicates any chemical addition
On-off operation of feed pump	Uniform sludge cake production	Simple and effective	Increases probability of plugged feed lines

Other control strategies that may be suitable and are currently being studied include control of the rotation speed and the vacuum of the filter, controlling the feed rate and addition of the coagulant based on the moisture content of the filtered sludge, and control of the feed of the iron coagulant based on the pH of the sludge conditioning tank.

INCINERATION

Sludge incineration is a combustion process that reduces the sludge volume and produces a sterile and easily disposed-of ash. Numerous proprietary incineration systems and furnaces, each with its own control scheme, are commercially available today.

This section touches on the important unit operations and processes that are automatically controlled in most incineration systems. For a more detailed examination, the reader should consult the literature recommended by relevant manufacturers.

Control of an incinerator and associated air pollution control devices can be classified by the following broad areas:

- Fuel feed
- Auxiliary fuel feed
- Combustion air
- Air pollution control.

In this instance the fuel would be raw and secondary waste sludge but, depending on its treatment, drying (or other processing) may be necessary prior to incineration. Control of the incinerator proper, however, entails control of the following parameters:

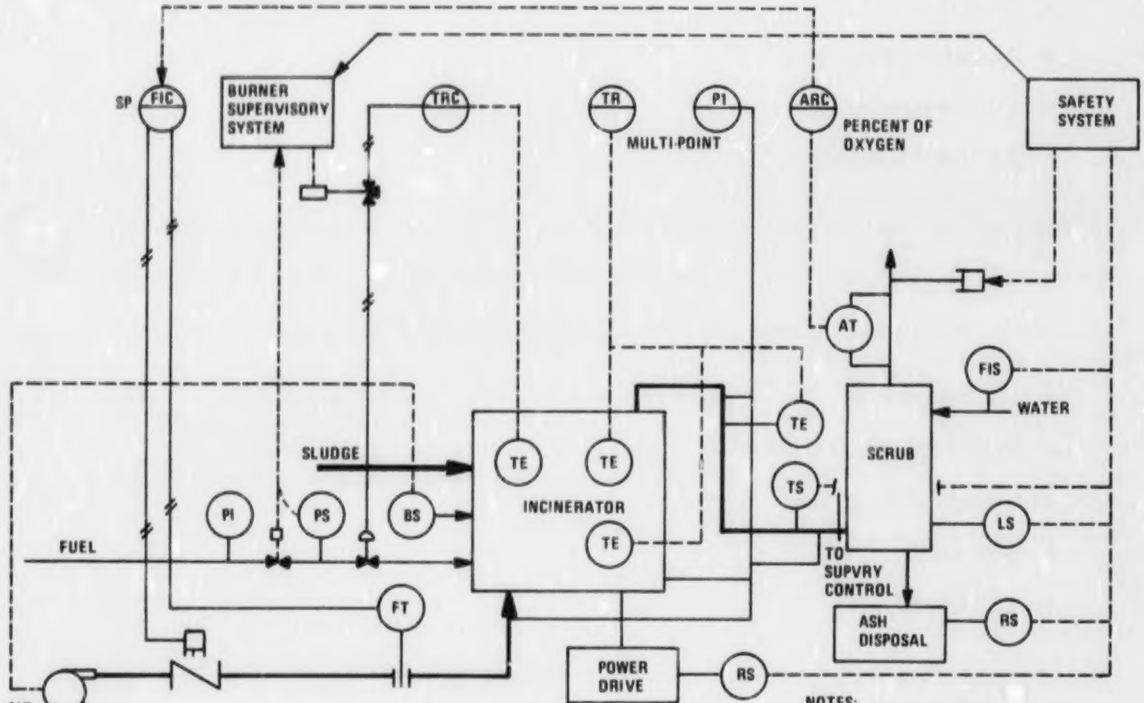
- Fuel feed rate
- Auxiliary fuel feed rate
- Grate airflow rate
- Top airflow rate
- Ash chamber.

The instrumentation associated with air pollution control devices is unique to each proprietary design. In general, though, it monitors and controls the following elements:

- Smoke density
- Water flow to scrubbers, etc.
- Stack gas composition (e.g., O_2).

Starting and shutdown sequences are complex and usually involve an analog programmer. Because of the potential hazards, elaborate safety shutdown systems are also included; these relate particularly to the auxiliary fuel burner equipment. Similar systems relate to specific types of air pollution control equipment in order to prevent equipment damage.

Figure 31 shows a typical set of control loops associated with commercially available incineration systems. The only control involving the operation of the incinerator is temperature. All other control loops relate to safety shutoffs, alarm systems and air pollution control. These control systems are supplied as part of the incinerator package.


SP
FIC
BURNER SUPERVISORY SYSTEM
TRC
TR
PI
ARC
PERCENT OF OXYGEN
MULTI-POINT
SAFETY SYSTEM
AT
FIS
WATER
TE
SCRUB
TS
SLUDGE
TE
TE
INCINERATOR
PI
PS
BS
TE
LS
FUEL
TO SUPVRY CONTROL
FT
ASH DISPOSAL
RS
POWER DRIVE
RS
AIR

NOTES:

1) Not shown: alarm system.
2) Starting circuit also requires full purge.
3) Safety circuit shuts down incinerator on:
    - Loss of main air
    - Loss of power drive
    - Loss of cooling water
    - Loss of fuel pressure
    - Loss of flame
    - Overtemperature
    - Overpressure.

Figure 31. Control system for a sludge incinerator.

NEUTRALIZATION OF ACIDS AND BASES

Control of the neutralization of acids and bases is a very comprehensive subject. Satisfactory control requires not only attention to the instruments and valves, but also to process chemistry, piping, mixing, and vessel design. As a result, this treatise can summarize only the recommended practices in all these areas.

General Chemistry and Reagents

The relationship between the amount of reagent added to a sample (or the flow of reagent added to a stream) and the resulting pH is known as a "titration curve" (Figure 32). The shape of this curve determines how well the effluent pH can be controlled, and the effort necessary to control it within specified limits. The shape of each curve is related to the strength of the acids and bases in the waste and the reagent, as well as their relative concentrations. References which detail acid-base chemistry as related to pH control are listed in the bibliography, particularly reference 15.

Weak acids and bases are especially helpful in buffering the pH of wastewater. The most common and useful natural buffers are the carbonates that occur in most water supplies. Carbon dioxide dissociates water weakly to form hydrogen and bicarbonate ions:

$$\frac{[H^+]\,[HCO_3^-]}{[CO_2]} = 10^{-6.35} \qquad (23)$$

The bicarbonate ion further dissociates into hydrogen and carbonate ions:

$$\frac{[H^+]\,[CO_3^{2-}]}{[HCO_3^-]} = 10^{-10.25} \qquad (24)$$

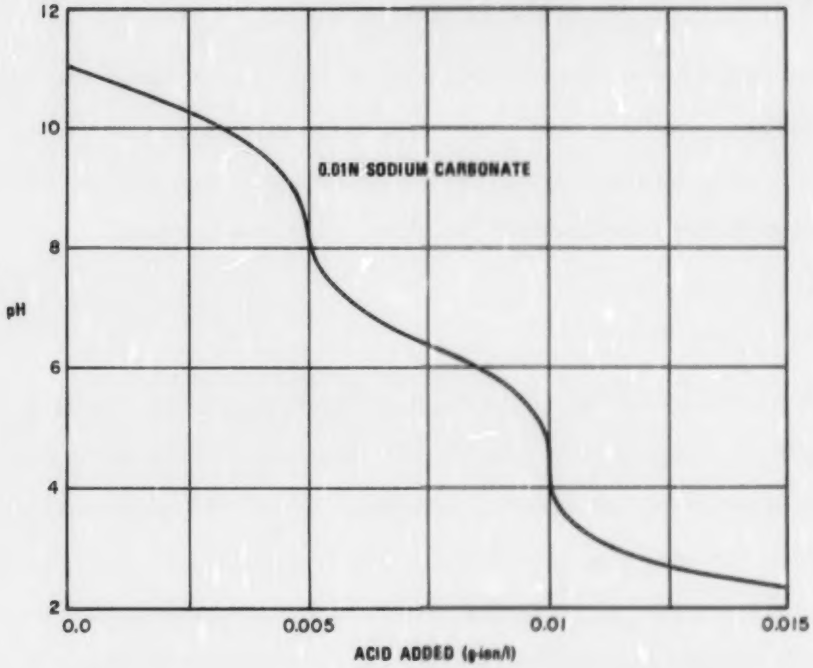


Figure 32. Carbonates provide valuable buffering in range of pH 6 to 8.

This double dissociation gives a titration curve with two intermediate plateaus in the vicinity of pH 6.35 and 10.25. Figure 32 illustrates this for a 0.01N carbonate solution titrated with a strong acid.

Most water supplies contain a measurable amount of alkalinity, which is expressed as ppm of $CaCO_3$. For waters of pH below 8, all of the reported alkalinity exists as the bicarbonate ion. Alkalinity values are, therefore, directly convertible to bicarbonate normality. The buffering effect of alkalinity in the vicinity of pH 7 can then be estimated as a function of alkalinity.

The choice of a reagent for neutralizing a waste should be based on ease of handling and effluent quality considerations, as well as cost and availability. Among the bases, lime is by far the easiest to handle, being available dry in bags for small plants, and in trucks or railroad cars for large consumers. It is also available in a 35% slurry in tank trucks as a byproduct of acetylene manufacture.

Due to the limited solubility of $Ca(OH)_2$ (1.16 g/l at 25°C), the maximum pH of a lime solution is only 12.5. Consequently, lime does not represent the hazard to workers or equipment that caustic does. However, it must be used as a slurry, which complicates mixing and metering.

Lime is available in two chemical forms: 1) high-calcium lime (93 to 98% CaO), and 2) dolomitic lime (55 to 58% CaO; 37 to 41% MgO). The latter is not generally recommended for effluent neutralization due to the very low solubility of the MgO and its sluggish reaction rate (16). Even high-calcium lime presents a residence-time problem due to its limited rate of reaction. Other alkaline reagents are limestone ($CaCO_3$), caustic soda (NaOH) and soda ash (Na_2CO_3).

The most common acidic reagent is sulfuric acid (H_2SO_4). Its principal advantages are its moderate cost and low corrosiveness in concentrated solution. Mild-steel vessels and pipe may be used to carry the acid. Other acidic reagents are hydrochloric acid (HCl) and carbon dioxide (CO_2)

Dynamic Response

Effective control of effluent pH depends on the dynamic response of the system and measuring device to a change in reagent flow. The best control will be achieved when the deadtime or delay in that response is minimized. All other things being equal, control effectiveness varies inversely with the square of the deadtime. Effective mixing is one method of reducing deadtime.

Figure 33 depicts a properly designed neutralization vessel. Blending is achieved by dropping the reagent directly into the influent as it falls into the vessel. The turbulence imparted by the fall provides the motive force. Concentrated reagents—especially sulfuric acid—evolve heat that is localized at the point of blending. Dropping the reagent into the open stream of influent (as in Figure 33) avoids the severe corrosion and boiling that would otherwise occur if the two streams met in a pipe.

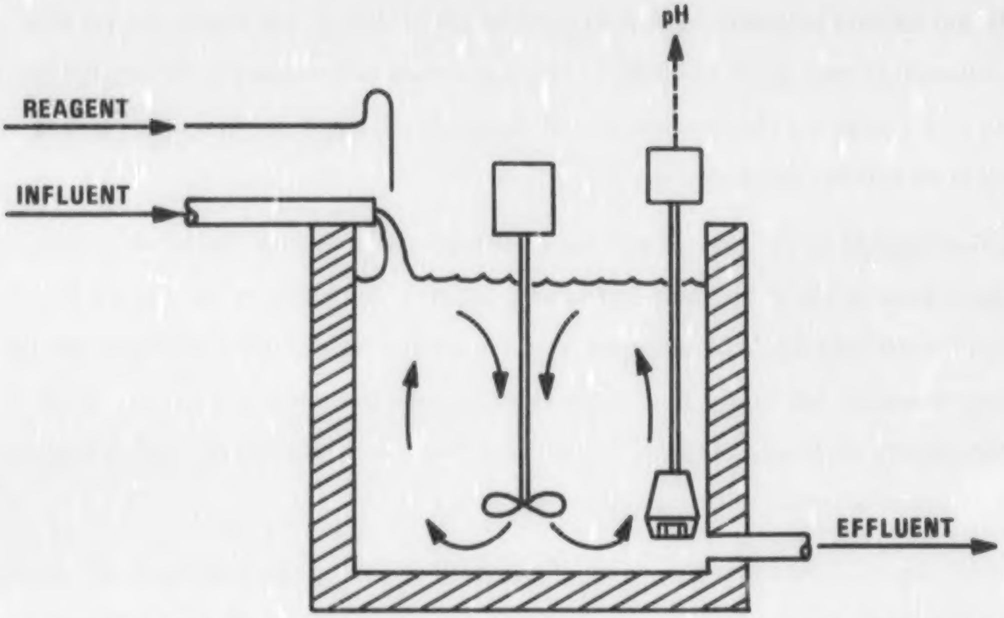


Figure 33. This configuration combines a thorough blending with fast response of pH to reagent flow.

The residence time in a vessel is its active volume, divided by the effluent flow rate. To make use of the entire volume, entry and exit should be diametrically opposite each other. The measuring electrodes should be located at the exit so that their potential truly represents the effluent pH. The agitator provides backmixing by recirculating the neutral effluent back to the feed point. At the same time, it pumps the thus-diluted feed to the point of discharge (and measurement) in a fraction of the actual residence time. Deadtime, as established by tank geometry between the time reagent is added and the measurement system senses the change, has been determined theoretically (15). It is half the tank volume, divided by the agitator pumping rate for cubic vessels laid out as in Figure 33. Verification is provided by Hoyle (17).

Hoyle (17) also demonstrates how feed entering one side of the vessel at the bottom requires twice as much time to reach the surface at the other side than if surface feeding is used. The reason for the difference is the downward circulation of the agitator, which forces the flow upward along the walls. Feed introduced near the bottom must flow upward along the wall, then downward along the shaft, and finally upward along the opposite wall before it can leave.

Propellers or axial turbines are recommended for mixing because they impart high turbulence. They are also less expensive than the slower radial turbines since they require no speed reduction. Shinskey (15) recommends about 2.5 HP/1000 gallons of vessel volume below 1000 gallons, 1.8 between 1000 and 10,000 gallons. Hoyle and others (17 and 18) support these recommendations.

Vortex formation must be avoided to properly utilize agitator horsepower. In a small vessel (100 gallons or less), off-center or off-vertical mounting of the agitator may suffice. In larger cylindrical vessels, vertical baffles are necessary. In cubic tanks, the corners tend to break up vortices although, in very large vessels, baffles may also be required. Influent should not be introduced into corners, nor effluent removed from them (especially if the vessel is baffled), as this adds deadtime to the response.

Deadtime is least when the depth, length, and width or diameter are equal.

Acid-base reactions are generally instantaneous, but occasionally a lag is encountered in ionizing the reagent. The limiting factor is the rate of solution of relatively insoluble reagents such as lime, limestone, and carbon dioxide. Particle size, purity, velocity, and pH all determine the rate of solution.

If insufficient residence time is allowed for the reaction to go to completion, the pH of the effluent at the control point may not be its final pH. If, for example, lime were added to a pH 2 influent to control the effluent at pH 7, its value could eventually reach 9 at some point downstream. A readjustment of the control point to pH 5 could yield a final pH of 7, but only at a constant load. If the influent pH were to increase to 5 at some later time, no lime would be added, and the final effluent pH would also be 5.

To minimize the difference between the controlled and final pH, residence time in a vessel using lime should be 15 minutes or more. The shorter the residence time, the greater these two values will differ. Shinskey (15) concludes that the slowest reacting component in lime reagent is the 1 to 2% $CaCO_3$ that is always present. Increasing the pH control point inhibits $CaCO_3$ solution and thereby yields an ostensibly faster reaction.

Although it may be possible to readjust the control point based on reagent flow in order to yield a constant final pH, this has not yet been done in practice. However, such a technique would apply to those processes with insufficient residence time to yield a stable effluent pH. Another possible solution to this problem is given in a later section.

Concentrated soluble reagents are used at very low flowrates requiring small metering pumps or control valves with small orifices. Dynamic response of the reagent flow to changes in the valve or pump control signal can be delayed if the piping downstream is allowed to drain freely. The loop seal in Figure 33 is used to prevent this from happening. The only section of line that drains is downstream of the loop seal; the seal, therefore, should be located at the point of entry into the vessel, whereas the valve or pump may be located at any convenient place.

Suspended reagents should be continuously recirculated back to the slurry tank by a centrifugal pump. The circulation loop should pass as closely as practicable to the neutralization vessel, so that the slurry may drop directly into it. Figure 34 shows that the spur from the circulating loop should proceed uphill into the horizontally mounted valve, and then downhill into the neutralization tank. In this configuration, the valve will be least likely to plug since the solids will drain away in both directions when the valve is closed. Flushing the lime valve or downstream piping continuously with water is not recommended, since its carbonate content (alkalinity) can form a scale of $CaCO_3$ where it meets the reagent.

The dry reagent may be metered directly onto the surface of the neutralization vessel, although some delay associated with dissolution will be encountered. Often the slurry is mixed in a separate tank that overflows into the neutralization vessel. Both the dry feed rate and the water flow into this slurry tank must be manipulated by the effluent

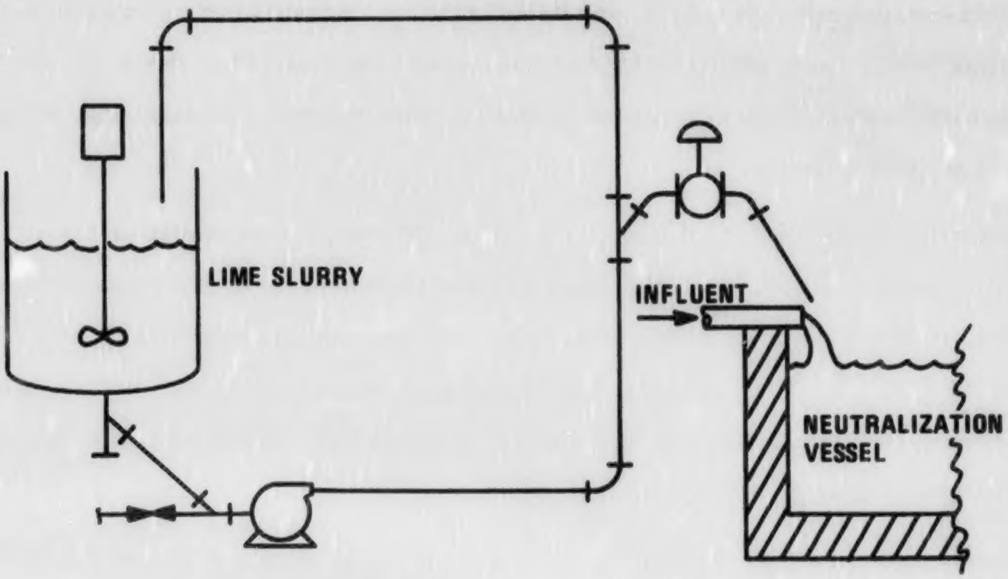


Figure 34. Slurry control valve should be mounted so that solids will settle away from it.

pH controller, or the slurry concentration will vary. If the water flows at a constant rate, changing the dry feed rate will affect the effluent pH only gradually because the concentration of the slurry must change first. With this added delay, the pH controller cannot keep up with even moderate variations in influent conditions.

More than one vessel may be desirable if effluent pH is to be controlled within the limits of 6 to 9 or better at all times. There are several reasons for this:

- Without heavy buffering, pH in the neutral range is so sensitive to reagent addition that rapid fluctuations and even cycling are often unavoidable.
- A sizable deviation from setpoint is usually needed before the controller can adjust reagent delivery to balance a large change in load.
- Temporary overloads are not uncommon, because process vessels are periodically emptied or cleaned, industrial chemicals are dumped into a municipal sewer system, etc.

An additional downstream vessel can attenuate the cycling and brief excursions in pH resulting from a rapid change in load or momentary overload (19). However, another pH measurement must be made at its discharge, with recording or alarming, although control is applied upstream.

An additional upstream vessel (i.e., one without pH control equipment) cannot attenuate cycling in controlled pH, but it can absorb load changes and temporary overloads as readily as a downstream tank. The upstream tank has the further advantage of providing an opportunity for acid and basic wastes to neutralize one another, thereby saving reagent. Plants treating both types of waste should, therefore, have an additional capacity upstream of the neutralization vessel.

Protection against an overload and component failure can also be provided by either an upstream or downstream vessel. Protection requires the coordination of three functions:

- Effluent discharge must be stopped when pH deviates beyond limits.
- Off-spec effluent must be recycled for additional treatment.
- Sufficient capacity for accumulating influent must be available to allow sufficient time for normal operation to be restored.

Figure 35 illustrates a two-vessel system that incorporates the protective features specified above. The two may be physically separate, or the neutralization vessel may simply be a partitioned corner of the larger tank. The submersible pump discharges or recycles effluent, depending on the position of the three-way valve. The valve is actuated by an alarm on effluent pH, allowing discharge only when applicable specifications are satisfied. Returning the recycle stream back to the point of influent entry provides some degree of mixing in the surge tank.

To avoid completely emptying the neutralization tank, a switch should deenergize the pump on low level. With the pump off, reagent flow may be shut off by suitable logic to avoid any waste. At the same time, the pH controller should be transferred to

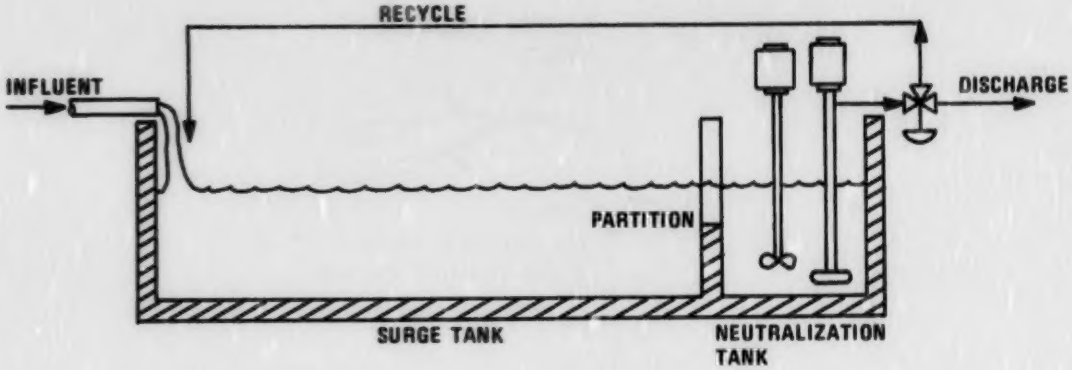


Figure 35. Three-way valve is actuated by an alarm on effluent pH, preventing discharge of off-limit materials.

manual; otherwise, a large transient will be sustained when the flow begins again. The agitator may be deenergized with the pump to conserve power. A system like this is described in greater detail in Shinskey (20, p 57).

Measurement

The measurement of pH has had a reputation of low reliability and high maintenance. The glass measuring electrode has an extremely high impedance (>100 megohms), and the circuit is therefore sensitive to electrical leakage. The reference electrode must maintain a liquid junction with the process fluid and, consequently, is subject to contamination. Being immersed in the process fluid also presents the possibility of fouling. However, the major factor in many failures is undoubtedly a lack of understanding of the principles of pH measurement.

The hydrogen-ion-sensitive portion of a pH electrode is a thin glass membrane (Figure 36). Behind this membrane is a solution buffered at pH 7. A new glass electrode requires an hour or more to hydrate, then the membrane develops a potential that is proportional to the difference between solution pH and the buffer at pH 7. To complete the circuit a reference electrode is necessary. The most common reference electrode used in the U.S. is a silver wire coated with silver chloride and immersed

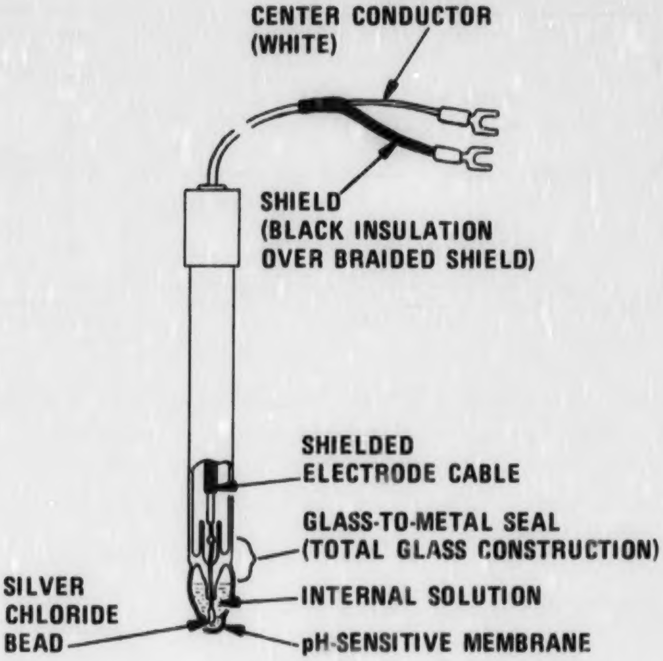

CENTER CONDUCTOR
(WHITE)
SHIELD
(BLACK INSULATION
OVER BRAIDED SHIELD)
SHIELDED
ELECTRODE CABLE
GLASS-TO-METAL SEAL
(TOTAL GLASS CONSTRUCTION)
INTERNAL SOLUTION
SILVER
CHLORIDE
BEAD
pH-SENSITIVE MEMBRANE

Figure 36. Construction details of a measuring electrode.

in a KCl solution saturated with AgCl. It is intended to develop a potential identical to the internal cell of the glass electrode.

Connection with the process occurs through a liquid junction at the tip of the electrode. A "flowing" electrode contains a reservoir of KCl electrolyte, which flows through a small orifice or porous plug at typically 1 ml/day. A flowing electrode is shown in Figure 37. Care must be taken to maintain this flow at all times, or the junction may become contaminated and cause an error in voltage. Temperatures below 19°C will cause both KCl and AgCl to crystallize from a 4.0M solution, tending to plug the liquid junction.

Nonflowing electrodes are completely filled with a saturated solution or gel of KCl and AgCl. They require neither reservoirs nor pressurization, thereby eliminating many maintenance and potential problems. Their accuracy is quite adequate for most waste-neutralization systems. Their life is limited, however, in that some seepage of electrolyte does occur, with a gradual contamination developing. Process solutions

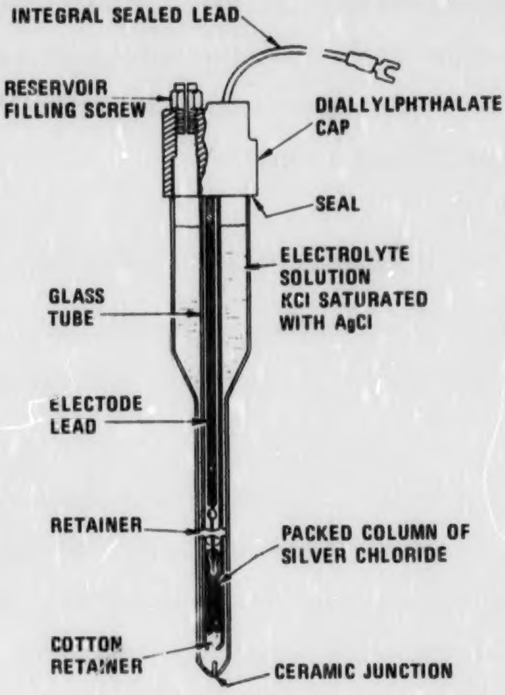

INTEGRAL SEALED LEAD
RESERVOIR FILLING SCREW
DIALLYLPHTHALATE CAP
SEAL
ELECTROLYTE SOLUTION KCl SATURATED WITH AgCl
GLASS TUBE
ELECTODE LEAD
RETAINER
PACKED COLUMN OF SILVER CHLORIDE
COTTON RETAINER
CERAMIC JUNCTION

Figure 37. Construction details of a flowing reference electrode (21).

containing sulfides or other ions with a great affinity for silver usually bring about more rapid deterioration.

Because the glass and reference electrodes have similar internal cells, no temperature error exists at pH 7 (0 mV), since waste neutralization is concerned primarily with accuracy in the pH 6 to 8 range, where the error is minimal. If temperature compensation is required it is readily available by inserting a resistance temperature detector into the electrode chamber. This resistance varies the gain of the pH amplifier.

In industrial installations the solution being treated is grounded through vessels, piping, etc. Because its ground potential may differ from that at the pH meter, ground currents could flow through the electrodes and cable. These currents are most likely to seek the path of lowest resistance; in the absence of a better conductor, this would be through the reference electrode. Although the ground currents tend to be ac, they can be rectified by the reference cell, producing a dc millivolt error. A solution ground wire is recommended to carry these ground currents from the solution to the instrument ground through a capacitor.

Possible Causes for Failure of Electrodes and Assemblies

In the event of failure of the electrodes or assemblies the operator will observe one of these possible symptoms: insensitive electrode, calibration drift, or slow electrode response.

When a pH measurement seems unusually constant compared to its past behavior, a malfunction is to be suspected. The usual diagnosis for this failure is electrical leakage which is caused by liquid leakage into the assembly. Field experience with submersible electrode assemblies has demonstrated how difficult it is to keep moisture out. Process solution could enter the lower chamber through an ineffective seal, or moist air could be drawn into the terminal box. Extreme variations in ambient conditions are probably responsible for most of the trouble. A reduction in ambient temperature will cause a partial vacuum within the assembly, along with condensation of moisture. If a leak does exist, more moist air will be ingested, with further condensation. Since moisture can enter but not escape, it accumulates. The moisture changes the resistance of the cables and terminal strips significantly, so as to cause a zero cell potential (pH = 7.0).

Packets of silica gel have been used to dry the air inside the assembly, but these are effective only in the absence of leaks. If a leak exists, the packets accumulate the moisture and short-circuit any terminals they touch.

Submersible assemblies (like the one shown in Figure 38) are most susceptible to leakage. Their superstructure is often at a very low or widely varying ambient temperture, while the submerged portion is at a solution temperature. The leakage of either solution or atmospheric moisture into the assembly can be eliminated with the

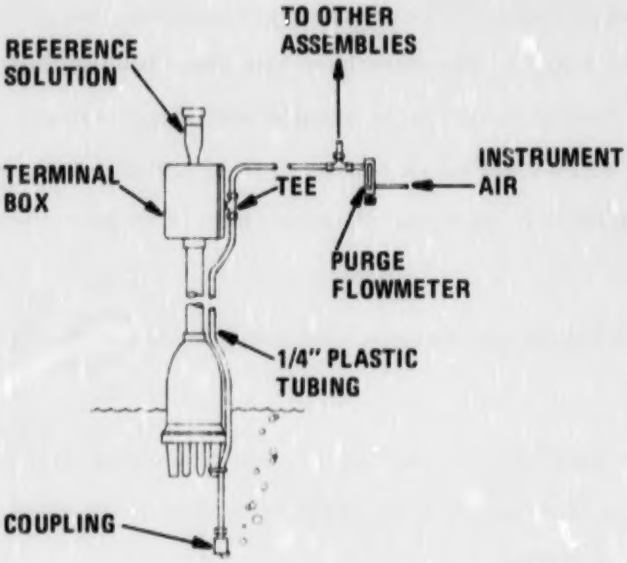


Figure 38. An air purge of submersible electrodes can eliminate leakage problems altogether.

air purge shown in Figure 38. The bubbler hanging below the electrodes will build a back pressure of air equivalent to the liquid depth. Then any penetration will allow only air to leak outward, rather than moisture to leak inward.

Instrument air is typically dried to, at most, a dewpoint of 4.5°C at 690 kPa (40°F at 100 psig), which corresponds to -20.5°C (-5°F) at an atmospheric pressure. Thus, condensation resulting from this source is not to be expected.

When the pH measurement is responsive, yet its calibration seems to drift from day to day, the reference electrode is suspect. The most probable cause is a plugged liquid junction, probably caused by the backflow of process fluid into the electrode, or by the precipitation of silver chloride from the reference solution. This contamination with the process solution will develop a variable electrode potential.

If backflow of process fluid appears suspect, then the solution is to relocate the following reference electrode within the reaction vessel so that proper flow is assured. A flowing reference electrode requires about 3 feet of head above the vessel liquid level to maintain a proper flow. If the vessel is pressurized, the electrode must be pressurized that much higher. (Vendors have kits available for this purpose.) On the other hand, if a flowing electrode is plugged with silver chloride, the electrode, its tubing, and reservoir should all be flushed free of crystals and refilled with 1.0 Molar solution. Restandardization of the pH meter may be required for a few days following the change.

A contaminated nonflowing reference electrode cannot ordinarily be repaired, but must be replaced.

Calibration drift can also be caused by an open solution ground or defective ground capacitor. It may also appear in the first hour after a dry glass electrode is installed, as the membrane hydrates.

Slow response of the electrode is another problem and the distinction between insensitivity and slow response is important. When the electrode assembly is moved from one solution to another, a failure to indicate the proper change in pH is considered insensitivity. If the change is indicated but only after several minutes have elapsed, sensitivity is there but response is slow.

In a control loop, insensitivity appears as a constant pH, while slow response causes oscillation of an increasingly or abnormally long period. Its primary cause is a film covering the glass membrane, which does not restrict the electron flow by electrical impedance but, rather, insulates the electrode from the varying composition of the solution. The ion flow from the solution to the electrode is impeded by the coating.

Coatings can develop from a variety of sources. Precipitation of such solids as $CaSO_4$, $CaCO_3$, and the hydroxides of heavy metals causes fouling. Greases, fats, and polymers already in the waste can also cause problems. Even a thin film of these substances may delay response significantly. Usually flowing reference electrodes are not susceptible to fouling, but glass and nonflowing electrodes are.

Installation and Maintenance

The pH meter or pH-to-current converter should first be calibrated with a portable potentiometer or electrode simulator connected in place of the electrodes. Only after the instrument is calibrated to the millivolt range corresponding to the desired pH scale should the electrodes be connected. A new glass electrode should be given an hour or more to hydrate and after that the electrode should not be allowed to dry out while not in use.

Standardization should always be made with the electrodes in a buffer solution of known composition. Distilled water should not be used for standardization because it is too easily contaminated. The pH meter allows a one-point calibration only, but always use two different buffers to confirm electrode sensitivity.

Whenever moisture is found in an electrode assembly or junction box, it should be removed with acetone, methanol, or a similar water-miscible, volatile solvent, followed by air-drying.

Coatings on the electrodes may be removed by periodic immersion of the entire assembly into a washing solution. Water-miscible solvents are suggested for greases and fats. A 10% (or thereabouts) solution of H_2SO_4 or HCl is extremely effective in removing salts and hydroxides.

When fouling is rapid, an ultrasonic cleaner (available from most instrument vendors) is recommended. This has been proven effective against clay, hydroxides, and calcium sulfate and phosphate precipitates. They do not seem to be effective on elastic polymers such as latex.

Mechanical cleaning is not recommended in that it is not as effective as the above methods in removing thin or hard films, and the probability of breakage is increased.

The presence of carbonates in alkaline waters reduces the sensitivity of the pH of water to the addition of strong acids and bases. An alkaline water containing 100 ppm

$CaCO_3$ equivalent at pH 8 may contain only 2 ppm when the pH is reduced to 5 or less. Below this level, virtually all bicarbonate ions are converted to carbon dioxide, which is then lost to the atmosphere. Neutralization with a hydroxide cannot restore this lost buffering capacity.

In the absence of buffers, the sensitivity of pH to reagent addition may be quite high. This means that a very small error in the ratio of reagent to influent flow can change the effluent pH significantly. For example, to provide effective control within a pH range of 6 to 8, the system would tolerate a 10% mismatch in terms of chemical equivalents between a basic reagent and an acid waste at pH 4, a 1% mismatch for a pH 3 waste, a 0.1% mismatch for a pH 2 waste, and so on. This example should indicate not only the sensitivity of the lightly buffered pH curve, but the effect of the initial concentration of the waste on control.

This extreme sensitivity makes control both difficult and necessary. The high gain of the pH curve requires a low controller gain if continuous cycling is to be avoided, but a low controller gain leaves the process susceptible to upsets in influent conditions. For these reasons, the process should be designed to be as controllable as practicable, following the guidelines given previously. Of equal importance is the application of control techniques and devices especially designed or selected for pH neutralization.

Final Control Elements

Final control elements such as metering pumps and control valves are discussed below.

Metering pumps with variable-speed drives are quite satisfactory for manipulating the flow of strong reagents. They are both linear and responsive, and are available in a sufficient variety of materials to deliver most commercial solutions used in neutralization. They can be sensitive to vapor lock or plugging with suspended solids.

Perhaps their most stringent limitation in waste neutralization is a rangeability of only about 20:1. Below about 5% of full speed, the motor tends to stall. When the reagent demand falls below this, limit cycling of the effluent pH will result. If an adjustable stroke is combined with adjustable speed, the rangeability is extended to a multiple of their individual rangeabilities. However, the gain of the pump then varies with stroke, affecting the performance of the control loop. Attempting to extend the rangeability by sequencing pumps of different size is also encumbered by the gain change with size. As a result, wide-range delivery is probably best accomplished with control valves.

Reagents are most commonly delivered through one or more control valves supplied from a head tank (in the case of solutions) or a recirculating loop (in the case of slurries). Valves are available with rangeabilities from 35:1 up to 100:1 or more, and in linear and equal-percentage (logarithmic) characteristics. For throttling slurries, ball valves are recommended—they have inherent equal-percentage characteristics.

A valve is said to be linear if a linear relationship exists between the stem position and the flow of liquid through the valve at a constant pressure. For equal-percentage valves, equal changes in valve stem position result in the same percentage change in liquid flow. In theory an equal-percentage valve will never shut off, but manufacturers provide modifications which allow the valve to shut off.

A linear characteristic is desirable for all applications except neutralization of a single, dominant weak acid or base. In that case, buffering varies with reagent demand. Fortunately, the variation in the gain of an equal-percentage valve in proportion to reagent flow will compensate this effect. In all other cases, the gain of the titration curve is either constant or not singularly related to reagent demand.

When the valve being used has the wrong characteristic for the process, characterization should be added. It may be implemented with a nonlinear positioner, a diode function generator, or a specially configured analog divider.

A valve positioner compares the control signal to the valve stem position, and acts on the motor to make them agree. Its most important function is to eliminate the hysteresis that is common to all actuators. Positioners should be used on all control valves used in waste-neutralization processes. Additionally, some positioners are available with contoured cams capable of positioning the stem as a nonlinear function of the control signal.

Figure 39 illustrates the nonlinear characteristic obtained by connecting the controller output (x) for an equal-percentage valve into both the numerator and denominator of a specially scaled divider. The divider will develop a biased signal which can be used to linearize an equal-percentage valve. When the divider output (y) is sent to the (linear) valve positioner, the overall relationship between the flow and controller output is nearly linear as shown. The divider is capable of higher gains than a contoured positioner and can, therefore, provide more exact linearization.

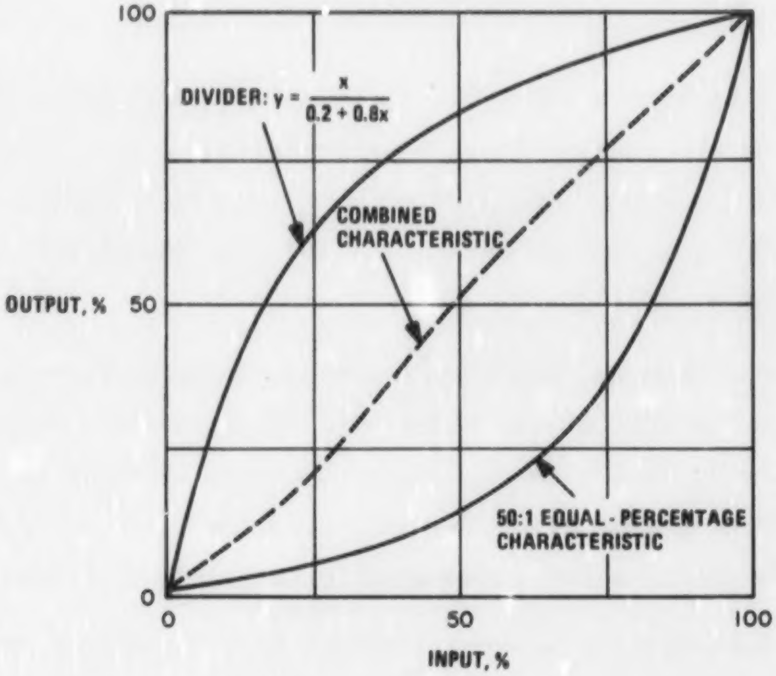


Figure 39. A properly scaled divider can effectively linearize an equal-percentage valve characteristic.

When both acid and basic reagents are used for neutralization, the valve positioners require different calibration. Both valves should be closed at 50% controller output, with the base valve fully open at 0% and the acid valve fully open at 100%. During an air-supply failure, however, both valves should close. Additionally, a solenoid valve on the air supply to the positioners can then be used to close both valves in case of a power failure or when the operator wishes to disable the system for maintenance, etc. If both valves are equal percentage, the required function for linearization takes the form of an "S," with the highest gain at midscale. This curve can best be obtained with a diode function generator.

Rangeability may be extended orders of magnitude by sequencing equal-percentage valves. Consider the need to manipulate flow over a 1000:1 range from 0.01 to 10 gpm. A 50:1 valve can throttle only to 0.2 gpm, but another 50:1 valve could cover the range from 0.01 to 0.5 gpm. Figure 40 shows the combined flow range on semilogarithmic coordinates.

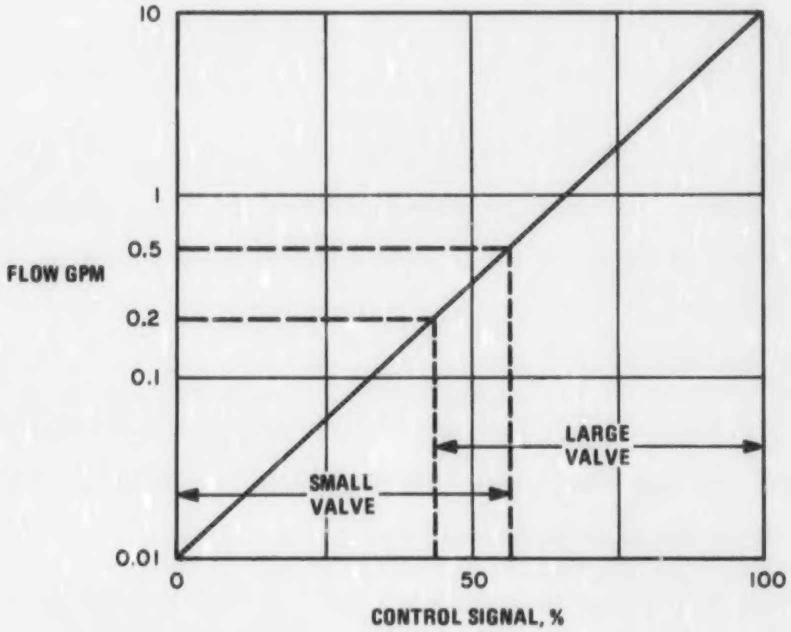


Figure 40. Two equal-percentage valves may be sequenced to act as a single wide-range valve.

The respective valve positioners must be calibrated to give full stem travel over the ranges of the control signal indicated on Figure 40. In addition, the valves must not be open at the same time. A pressure switch set to close when the control signal exceeds the full output of the smaller valve and to open below the minimum flow of the larger valve can actuate three-way solenoid valves to select which control valve will be open.

Feedback Control

The accuracy with which reagent flow must match influent demand places the burden of meeting effluent pH limits squarely on the feedback controller. Although feed-forward control may be used in some situations and may be helpful, it cannot approach the accuracy of reagent delivery needed. All of the foregoing recommendations regarding vessel layout, mixing, etc., are intended to make feedback control more effective.

An ideal feedback controller can be described by the following equation:

$$m = \frac{100}{P}\left(e + \frac{1}{R}\int e\,dt + D\frac{de}{dt}\right) \tag{25}$$

Output m is proportional to the deviation e between measurement and setpoint through the proportional band P, expressed in percent. The output is also related to the time integral of the deviation through the integral or reset time constant R. Finally, the output is also affected by the rate of change or time derivative of deviation through the derivative time constant D.

The primary controlling modes are proportional and integral. Derivative is helpful in accelerating response to upsets, but its effectiveness is limited by sensitivity to noise (uncontrollable rapid fluctuations) in the measurement signal.

Integral is necessary to reduce the deviation to zero and, the smaller the integral time, the faster the deviation will be reduced. R cannot be reduced below the deadtime in the loop, however, or an undamped oscillation will result. Theoretically, R and D can both be set equal to the deadtime times $2/\pi$, as developed by Shinskey (22). However, variations in the slope of the titration curve, reaction rate, and electrode response can lead to instability with such close settings. In practice, the two settings are best spaced 4:1 apart, such that R is $4/\pi$ times deadtime and D is deadtime divided by π. In situations where derivative is not or can not be used due to a high noise level, R ought to be increased to two deadtimes for stable response.

The proportional mode combines stability and responsiveness, but P must be set high enough to avoid continuous cycling. Most pH curves are sufficiently sensitive in the neutral range that stability requires a very wide proportional band, often exceeding 200 or 300%. As a result, the controller is not especially responsive to upsets, and large deviations tend to persist. This problem is due not only to the sensitivity of the pH curve, but also to its extreme nonlinearity. Small deviations from setpoint require the slightest corrective action, whereas even moderate deviations may require drastic changes in reagent delivery.

In cases where the proportional band cannot be set wide enough for stability due to its limited range, a limit cycle will appear. This cycle tends to be of a regular period and nearly uniform amplitude, usually—although not always—symmetrical about the setpoint. Any variation in its amplitude would indicate a change in the titration curve.

Adding a compensating nonlinear function to the controller tends to satisfy both the extreme proportional-band requirement and the nonlinearity of titration curves. Controllers that have simple nonlinear functions imposed on the deviation signal prior to action by the three conventional control modes are commercially available. The function consists essentially of three straight line segments that are symmetrical about zero deviation, as shown in Figure 41. The width of the gap is adjustable, as well as the gain within it, to permit matching to the titration curve; gain outside the gap is unity.

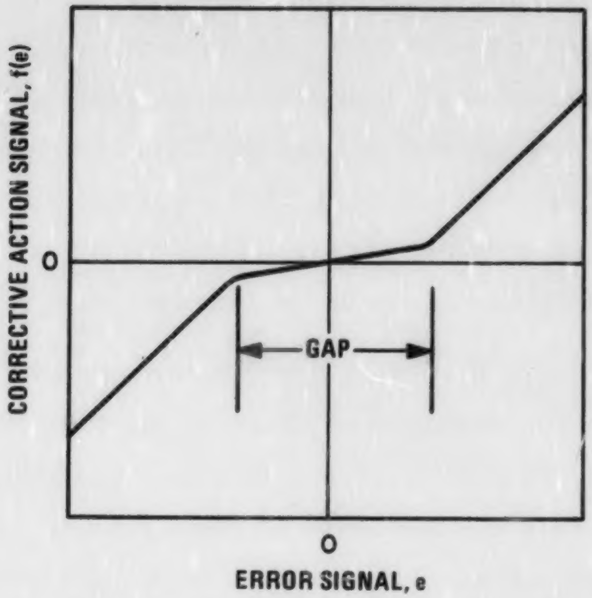


Figure 41. This nonlinear function greatly improves pH control.

In practice, the gap width is set to quench the limit cycle that would otherwise develop with a nominal proportional band of 100%. The proportional band is then set to recover rapidly from deviations exceeding the gap, but also to avoid limiting cycling outside the gap. The gain within the gap is set to avoid cycling within it. Too low a gain setting, however, may promote a slow cycle that would be equal in amplitude to the gap width, as explained by Shinskey (15).

A companion to the nonlinear controller is available, which automatically adjusts the gap width for applications where the titration curve varies extensively. This instrument is described by Shinskey (23). Called an adaptive controller, it senses an oscillatory or drifting condition of the pH loop and restores stable control.

The same pH deviation on which the nonlinear controller operates is sent to the adaptive controller. Its output is, in turn, connected to a remote gap-width adjustment in the nonlinear controller. Oscillation at or near the natural period of the pH loop will cause the adaptive controller to gradually expand the gap until stability is restored;

the gap will then remain at the value until another undesirable condition appears. When a steady or slowly drifting deviation develops, the adaptive controller will gradually narrow the gap until zero deviation is achieved.

The rate at which the gap is changed is based on the magnitude of the deviation and the integral time of the adaptive controller. A low-frequency gain adjustment is available to reduce the rate of gap closure compared to the rate of expansion. This additional parameter was found necessary for stability due to the extreme difficulty in controlling pH precisely at the setpoint. Considering that it is required only to control within a band (e.g., 6 to 8.5), small amounts of offset are tolerable, whereas small-amplitude cycling may not be.

A third adjustment tunes the adaptive controller to the natural period of the pH loop. It is important that the period be nearly constant when an adaptive controller of this type is used. Consequently, in cases where electrodes foul rapidly, continuous cleaning (as with an ultrasonic cleaner) should be provided.

The adaptive controller is especially useful when the influent is comprised of a multiplicity of wastes, including both strong and weak acids and bases. Batch operations, cleaning and rinsing, and periodic shutdown of parts of the production may cause these agents to come and go in varying proportions. As a result, the nonlinear controller may rarely be properly adjusted without automatic adaptation. Persistent limit cycling, during which acid and basic reagents are alternately added, can be avoided with adaptation. A properly adjusted nonlinear-adaptive control system requires only a few cycles to expand the gap to the point extinction.

Feedforward Control

Feedforward control is defined in Reference 24 as: "Control in which information concerning one or more conditions that can disturb the controlled variable is converted into corrective action to minimize deviations of the controlled variable."

With regard to waste neutralization, the "conditions" mentioned above are the demands for reagent needed to neutralize the influent. To achieve neutrality:

$$F_A N_A = F_B N_B \tag{26}$$

where F and N represent the flow rate and the chemical equivalents of acid (A) and base (B). Either may be the waste or the reagent. To apply feedforward control, Equation 26 is solved for the reagent flow in terms of the variable flow and equivalent concentration of the waste, as well as the fixed equivalent concentration of the reagent.

If the waste flow can vary rapidly but its composition cannot, then the feedforward system reduces to a simple flow ratio of reagent to influent, with the ratio adjusted by the effluent pH controller. This system requires a linear influent-flow measurement and a linear reagent-flow control loop or final element. The principal drawback to these linear systems is that their rangeability is limited to that of the final element and influent flowmeter.

When a metering pump is the final element, either of the two arrangements shown in Figure 42 may be used. If the pump has only variable speed or stroke (but not both), a multiplier is required to combine the influent-flow signal with feedback trim from effluent pH. If both stroke and speed are remotely manipulable, the feedforward signal may manipulate one while feedback adjusts the other. Their multiplication is inherent in that reagent flow is the product of stroke and speed.

The interlock between the reagent flow and the influent pump described previously is a flow-feedforward system in which only two values of flow are used. Regardless of the reagent flow at the time the pump is deenergized, it will drop to zero. When the pump is restarted, the reagent flow will return to its last value, which is the most probable new demand of the influent.

Plants wherein wastes are directed to the neutralization facilities under gravity head do not impose sudden changes in flow on those facilities. Influent flow cannot change

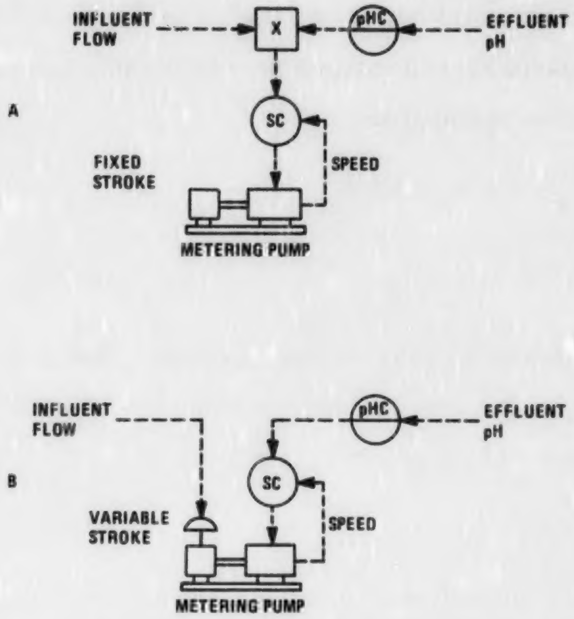

INFLUENT FLOW
X
pHC
EFFLUENT pH
A
SC
FIXED STROKE
SPEED
METERING PUMP
INFLUENT FLOW
pHC
EFFLUENT pH
B
SC
VARIABLE STROKE
SPEED
METERING PUMP

Figure 42. Flow-feedforward systems for fixed stroke (A) and variable stroke (B) metering pumps.

without first raising or lowering the level of liquid in the channels feeding the facility. The capacity of these channels, therefore, absorbs the flow variations effectively.

However, spills, rinses, and sudden discharges tend to be carried through the channels almost as if they rode a conveyor. Arriving at the neutralization facility only moderately diluted or blended with the remainder of the influent, they impose sudden reagent demands on the control system. An influent surge tank, as shown in Figure 35, is recommended to distribute these upsets. But where such a tank does not exist or can not be installed, feedforward control is the only choice remaining to diminish their effect on effluent pH.

Equation 26 could be solved directly by the control system in terms of reagent flow, if influent concentration were known directly. Analyzers that can make the determination of total acidity or alkalinity are available, but their dynamic response tends to be slower than the process itself. As a result, they are useful only in the isolated case of an especially slow feedback-loop response.

Influent pH is the measurement normally used for feedforward control (19). Converting Equation 26 to logarithms will indicate the relationship between reagent flow and influent pH, if ionization is complete.

$$\begin{aligned}\log F_A + \log N_A &= \log F_B + \log N_B \\ \log F_A - pH_A &= \log F_B - pH_B\end{aligned} \tag{27}$$

Equation 27 is implemented by using an equal-percentage (logarithmic) valve to relate reagent flow to influent pH. The schematic for such a system is shown in Figure 43. The actual equation solved by the system is:

$$s = m \pm K\,(pH - r) \tag{28}$$

where s is the signal to the valve(s), m is the feedback controller output, and K is a gain adjustment. The term "r" is a feedforward reference value that represents the influent pH at which there is no feedforward contribution to the valve position. The sign preceding K is positive for a basic influent and negative for an acidic influent.

For completely ionized influents, K is adjusted to produce a tenfold change in reagent flow per unit change in pH. [Its numerical value depends on the rangeability of the valve(s) and the span of the pH measurement.] Single weak acids and bases that change their pH only 0.5 unit per decade concentration would require K to be doubled.

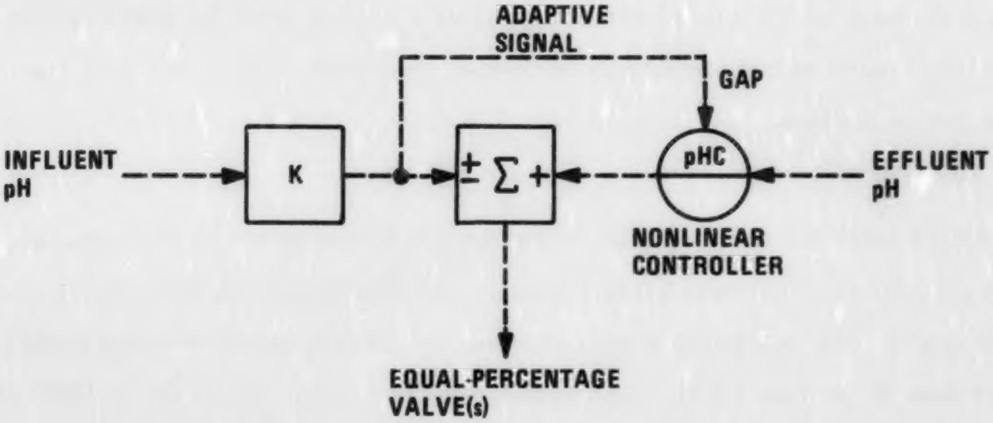


Figure 43. pH feedforward system is capable of a wide range through sequenced equal-percentage valves.

Mixtures of strong and weak components, however, vary the relationship between pH and concentration so widely (15) that a field adjustment to the best average value is the final recourse. This factor severely limits the improvement that feedforward control brings to the neutralization process (19).

When manipulating equal-percentage valves for pH control, the gain change caused by the valve characteristic adversely affects the stability of the feedback loop. To overcome this problem, the feedforward component of valve position [i.e., K(pH - r] is used to set the gap width of the nonlinear controller. Thus, when the influent pH is moderate (calling for little reagent), the gap will be narrow (compensating for a diminished valve gain). Figure 43 indicates this as an adaptive loop in that the controller is automatically adjusted as process conditions change.

At present, the hydraulic analog control has not been proven in practice. Its success is contingent on the satisfactory performance of certain hydraulic operations that still require some development. Nonetheless, it is included here for possible further examination by those who are confronted with a particularly unyielding neutralization problem.

The problem is the lack of sufficient residence time for neutralization with lime in a conduit having no backmixing. Figure 44 illustrates the addition of a small stirred

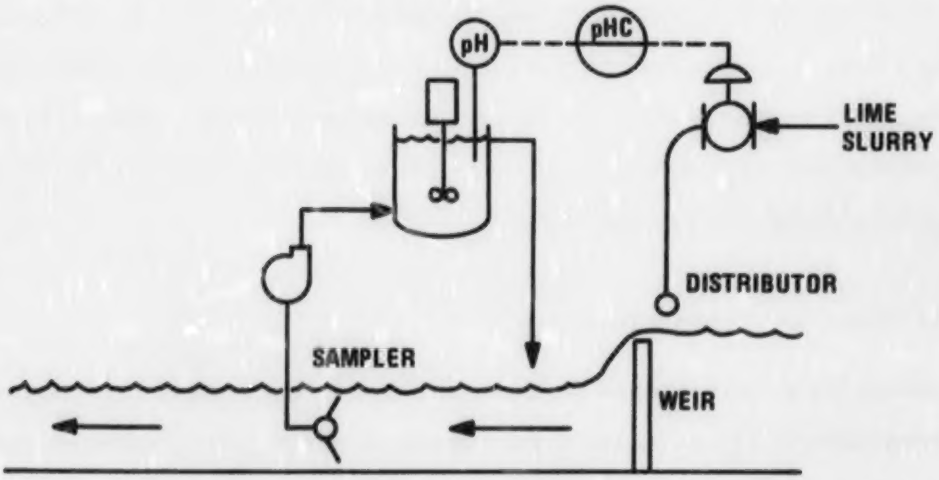


Figure 44. Small tank is a hydraulic analog of larger stream.

tank with sufficient residence time, to which a sample of the treated stream is sent. The pH in the tank is controlled by adding a reagent to the main stream.

Whether such a configuration is workable depends on uniformly distributing the reagent across the stream and obtaining a representative sample of the treated stream. Even a 1% variability in either of these functions could cause a 1 or 2 pH deviation in the final condition of the effluent. Distribution may be more difficult than sampling in that it must be effective over the entire range of reagent flow. The weir will help distribution, and downstream baffling should improve the blend. Locating the sample point too far downstream, however, will add to the deadtime of the control loop.

PHOSPHOROUS REMOVAL BY LIME TREATMENT

Owing to the increasing concern over the rate of eutrophication of our surface waters, many states have instituted standards for phosphorus removal. The areas most affected are wherever both industry and agriculture are concentrated along lakeshore and riverfront, particularly in areas bordering on the Great Lakes. Typical among standards instituted by these states is Indiana's Water Quality Standards SPC1R-3, requiring 80% removal or 1.0 mg/l phosphorus, whichever is more stringent.

In order to meet these or any standards, adequate measuring and controlling systems must be applied to the treatment process. One of the available phosphorus removal processes is the precipitation of calcium phosphate salts through addition of lime. The design aspects of this process are taken primarily from the "Process Design Manual for Phosphorus Removal" (25).

General Theory and Process Dynamics

The process design manual cites the three principal forms in which phosphorus may enter wastewaters: orthophosphate, polyphosphate, and organic phosphorus compounds. The organic phosphorus compounds largely break down during biological treatment into the orthophosphate. Polyphosphates come primarily from detergents and corrosion inhibitors, but they, too, ultimately hydrolyze into orthophosphates. Consequently, phosphorus control in wastewaters is centered around the removal of the orthophosphate ions by precipitation.

The solubility of salts is governed by an equilibrium between their ionic components. Consider the precipitation of calcium orthophosphate:

$$3Ca^{2+} + 2PO_4^{3-} \rightleftarrows Ca_3(PO_4)_2\downarrow \qquad (29)$$

When the product of the activities of the individual ions raised to the powers of their proportion in the solid reaches the solubility-product constant, the solution is saturated with respect to that solid:

$$\left[Ca^{2+}\right]^3 \left[PO_4^{3-}\right]^2 = 10^{-26} \qquad (30)$$

Bracketed terms refer to ionic activity in units of mols/liter. Under the dilute conditions usually found in wastewaters, activity and concentration are nearly identical. Therefore, the bracketed terms may be considered as concentration in mols/liter with little error in most cases.

Stoichiometric equations such as Equation 29 can be used to estimate the dose of calcium required to react with the orthophosphate. Solubility product equations such as 30 indicate the excess quantity of calcium needed above the stoichiometric demand to drive the precipitation reaction toward completion.

The multi-basic nature of the phosphate species complicates the picture, however. As a function of pH (to be described below), a hydrogen-phosphate ion forms, which may also be precipitated by calcium ions:

$$Ca^{2+} + HPO_4^{2-} \rightleftarrows CaHOP_4\downarrow \qquad (31)$$

The solubility of this salt is given as:

$$\left| Ca^{2+} \right| \left| HPO_4^{2-} \right| = 10^{-7} \qquad (32)$$

Figure 45 illustrates the theoretical solubility of phosphates in lime solutions, plotted as total dissolved P on a logarithmic scale against pH. Assuming an initial concentration of 30 mg/l P, no precipitation will result from lime addition until pH 6.5. If more phosphorus is present, precipitation will begin at a lower pH or, if less, at a higher pH. The curve representing mg/l P vs pH for $CaHPO_4$ precipitation is fixed; only the point of intersection with the initial concentration is dependent on that initial concentration.

As pH is raised further by lime addition, the rate of $CaHPO_4$ precipitation (the slope of the curve) diminishes. Above pH 8, enough PO_4^{3-} ions are formed to begin

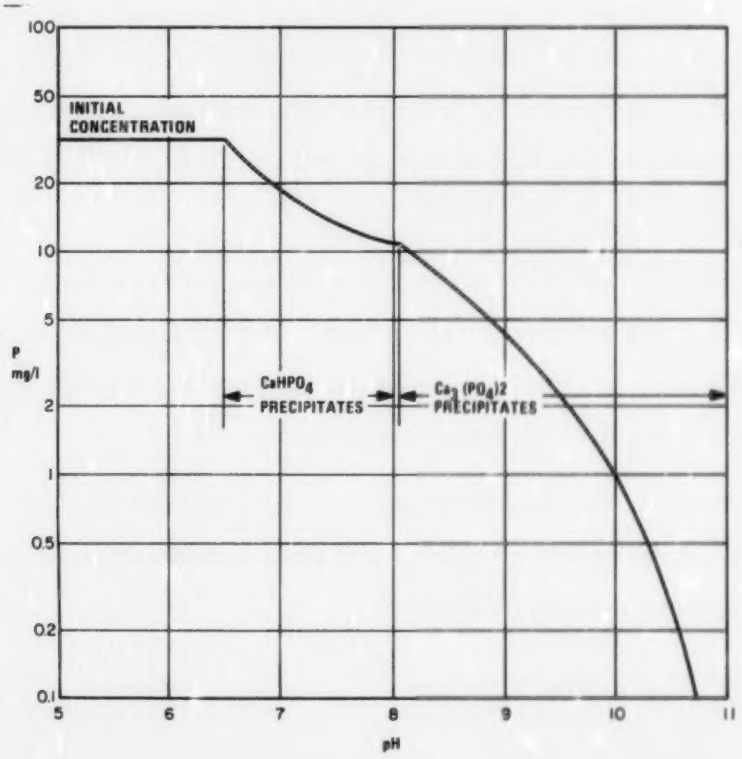


Figure 45. Theoretical solubility of phosphates in lime solutions as a function of pH.

precipitation of $Ca_3(PO_4)_2$. Butler (26) calculates the pH of a solution saturated with both $CaHPO_4$ and $Ca_3(PO_4)_2$ as 8.06.

As the pH increases above this point, $\left[PO_4^{3-}\right]$ concentration increases, which reduces $\left[Ca^{2+}\right]$ concentration. Consequently, $CaHPO_4$ no longer precipitates above pH 8.06. Lime addition above pH 10 begins increasing $\left[Ca^{2+}\right]$ again, causing total P to drop sharply.

However, Figure 45 describes a theoretical relationship for the solubility of only the orthophosphates. Other forms of phosphorus besides orthophosphate may or may not precipitate, resulting in a different level of total P. Albertson and Sherwood (27) have found that 9 mg/l PO_4 (which would be 3 mg/l P) at pH 9 and 4 mg/l PO_4 (1.3 mg/l P) at pH 10, are soluble after lime treatment of domestic sewage.

Figure 46 indicates lower solubilities of total P at pH 9 to 11 than theory indicates. All the points represent data taken from the EPA pilot plant at Lebanon, Ohio. The reduction in the phosphorus level below the solubility of calcium phosphates must be due to other reactions. The wide variation in the data points indicates the variability

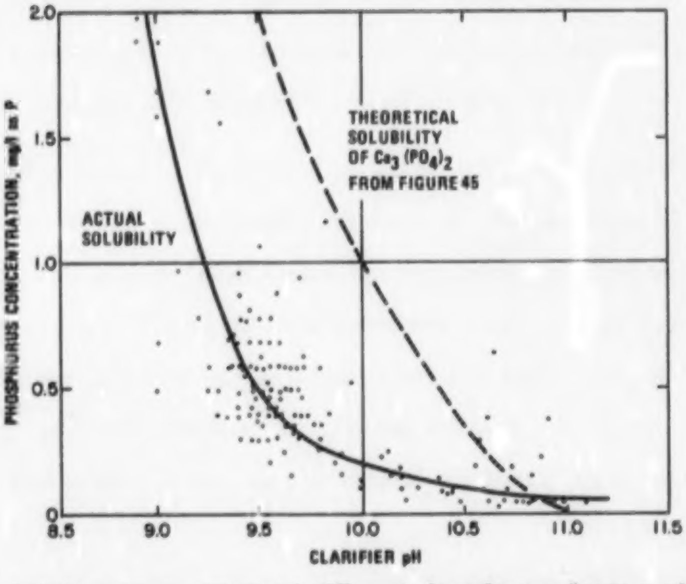


Figure 46. Actual measurement of effluent phosphorus from a single-stage lime treatment process as a function of pH, compared to theoretical relationship.

of these undefined reactions or the components in the water that determine their course. The disagreement between this data and that cited above by Albertson and Sherwood (27) indicates that other characteristics of the water or the treatment plant have some influence over the phosphorus level in the effluent.

The conclusions that can be drawn from the above information are these:

- pH control is necessary for effective phosphorus removal.
- Phosphorus concentration in the effluent cannot be correlated exactly with pH from one plant to another, nor even from time to time within the same plant.

It has been demonstrated that lime addition to a pH of 9.5 or 10 is necessary to reduce effluent phosphorus levels below 1.0 mg/l. If the effluent were to be discharged to a receiving body of water at these high pH levels, damage to aquatic life could result. Hence, the pH must be reduced below 8.5 (or whatever applicable standards require) by adding an acid of some type. Addition of CO_2 to lower the pH (recarbonation) is widely practiced.

If the wastewater alkalinity is already high (>150 mg/l $CaCO_3$), lime addition to pH 9.5 to 10 will precipitate enough $CaCO_3$ along with the phosphates to form a settleable floc. After clarification, the pH is adjusted to 8 or thereabouts by recarbonation. The data in Figure 46 came from a single-stage plant such as this.

With wastewaters of low alkalinity, lime addition to pH 9.5 will not precipitate enough $CaCO_3$ for good floc formation. In these situations, the pH must be raised by lime to 11 in order to precipitate $Mg(OH)_2$ and as much $CaCO_3$ as possible, along with the phosphates. Then a first-stage recarbonation to pH 10 precipitates more $CaCO_3$ to promote floc formation. After clarification, a second stage of recarbonation adjusts the pH to its final value of 8 or thereabouts. No precipitation takes place during the second stage, since the minimum solubility of calcium in a system saturated with CO_2 occurs about pH 10.

Lime treatment for phosphorus removal may be applied to the raw wastewater prior to biological treatment. CO_2 generated during biological treatment is absorbed, and may lower the effluent pH to acceptable levels, so that no neutralization is required. A pH of less than 9.5 is generally accepteable prior to biological treatment (25).

The success or failure of chemical waste-treatment processes hinges to a great extent on the accuracy and responsiveness of the chemical feeding system. This is particularly true when a slurry is being fed. If failures are frequent due to plugging, etc., controls are usually bypassed and feed rates left at some constant maintainable value. This practice results in the overfeeding and waste of chemicals, and usually creates other problems. But a well-designed feeding system can be both reliable and responsive, and will operate indefinitely under automatic control.

In tonnage quantities, pebble lime (CaO) is slaked with water to form hydrated lime [$Ca(OH)_2$] in an apparatus similar to that shown in Figure 47. Because the reaction is exothermic, heat is released as steam. Consequently, enough water must be supplied to make up the loss as well as provide the desired concentration of slurry.

In order to avoid having to adjust feeding the slaker over the entire range of lime rates demanded by the waste-treatment process, the slaker should be operated on-off. As the level in the slurry tank reaches a low point, the lime feeder and water flow are started at preset rates. When the level reaches the high limit, both are stopped.

Because the lime-slurry control valve may be only 1/2 inch in size, a vibrating screen should be used to take out the gravel always present in pebble lime. The lime slurry should be continuously circulated to avoid plugging. The control valve should be mounted at the highest point in its branch, so that solids will not settle there when the valve is closed. In addition, there should be no horizontal sections in the branch containing the valve. The valve should be mounted as close to the point of discharge into the reaction vessel as possible, to minimize any delay in response to the control signal. Avoid flushing continuously with water though, as $CaCO_3$ tends to form a scale wherever the water meets the lime.

The control valve should be a ball valve for rangeability and resistance to plugging. To minimize wear from sand in the slurry, the piping should be selected for a 5 psi

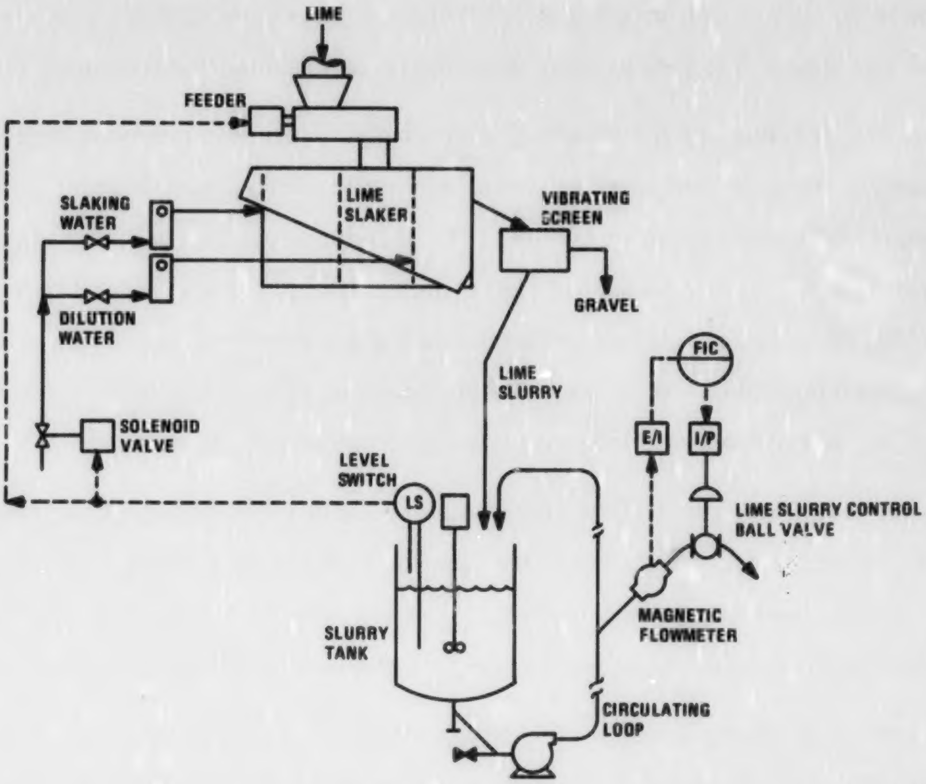

LIME
FEEDER
SLAKING WATER
LIME SLAKER
VIBRATING SCREEN
DILUTION WATER
GRAVEL
FIC
LIME SLURRY
E/I
I/P
SOLENOID VALVE
LEVEL SWITCH
LS
LIME SLURRY CONTROL BALL VALVE
SLURRY TANK
MAGNETIC FLOWMETER
CIRCULATING LOOP

Figure 47. Control of lime slaking.

or less drop across the valve. Since control ball valves are not available in sizes less than 1/2 inch, the slurry concentration should be selected with this in mind.

A magnetic flowmeter is recommended since it offers an unobstructed path and provides a linear measurement of flow. It must be mounted in a vertical or sloping upflow section, as shown in Figure 47. A flow controller closes the loop to the valve; however, it may in turn be set by a pH controller or other signal, as described later in this section.

Smaller plants will find hydrated lime in 50-lb bags easier to use. Typically, the dry lime is metered into a slurry tank, as shown in Figure 48. The slurry then overflows into the reaction vessel at a rate proportional to the flow of water into the slurry tank.

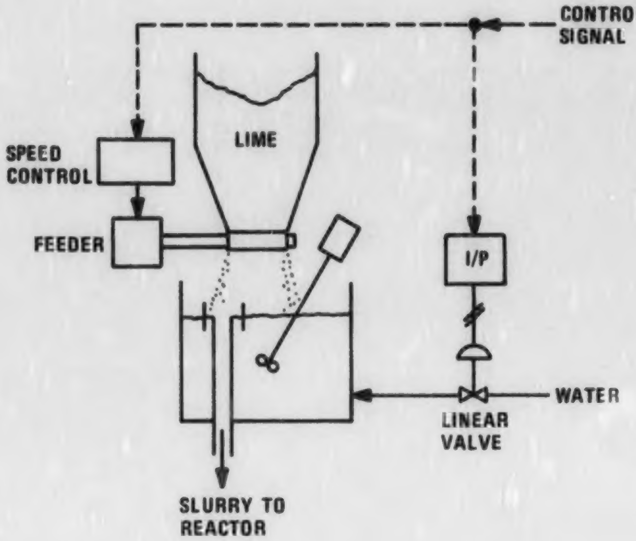


Figure 48. Control of dry lime feed.

To simplify the controls, some contractors fail to put a control valve on the water, leaving it to enter at a constant rate. This practice unfortunately creates a serious time lag; when the lime feeder is stopped, for example, the lime in the slurry continues to overflow until the tank is depleted. This renders control over the reaction vessel virtually impossible.

But if a linear control valve on the water is made to follow the same signal as the lime feeder, the concentration of the slurry will stay nearly constant. When the control signal changes, the rate of overflow will then change, and the reaction will be responsive and controllable.

Figure 49 compares the records of pH control of a neutralization vessel. In the top record, water into the slurry tank is flowing at a constant rate. In the lower record, water flow as well as lime feed rate is manipulated by the pH controller. The second pH measurement shown in both records represents the effluent from the clarifier downstream of the neutralization vessel. The superior degree of control achieved

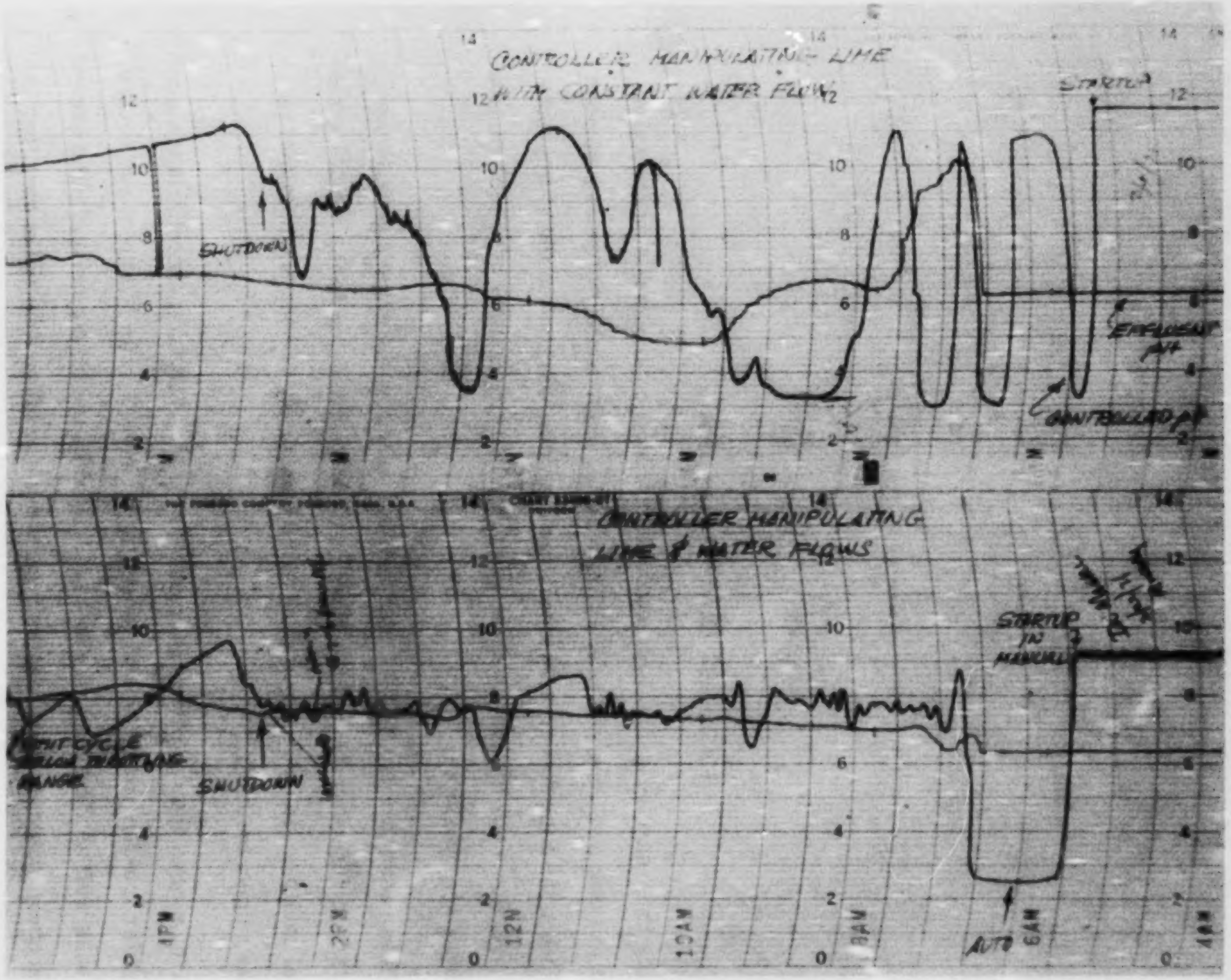


Figure 49. Upper record shows uncontrollable pH caused by water flowing at a constant rate into slurry tank; when water flow was manipulated with lime flow, lower record was produced.

with the latter system is obvious. (Note that time proceeds from right to left on the charts.)

Lime calcining or exhausts from combustion engines or submersible burners are economical sources of CO_2 for large users. If the source is dedicated CO_2 generation for recarbonation, then the control signal simply manipulates the fuel flow or burner mechanism. Since any combustion system has a low fuel limit to avoid flameout, the rangeability of these sources is not wide.

If the exhaust from a calciner or internal combustion engine is used as a CO_2 source, its availability is fixed by other considerations. Whatever CO_2 is not used for recarbonation must be vented to the atmosphere. The control signal would then manipulate dampers in the feed line and stack, opening one while closing the other. Since the CO_2 from this source may vary between 6 and 18%, automatic control from a continuous signal such as pH must be used, rather than simple pacing or flow control.

Carbon dioxide may be purchased as a liquid under pressure in cylinders. Feeding then requires vaporization and metering of the dry gas. For rates less than 1000 lb per day or so, enough heat is available from the surroundings for evaporation. Higher rates require a source of heat and temperature or pressure control.

Commercial feeding equipment is available from several vendors, who also furnish chlorinators. The feeding equipment usually has built-in metering components, so that flow is proportional to a standard control signal. The response of the process to control is usually satisfactory if the feeder is located not more than 100 feet from the recarbonation vessel.

Process Control

Since wastewater flow rates tend to fluctuate diurnally (if not more often), feeding chemicals at a constant rate cannot be satisfactory in the long term. If the composition of the waste and the chemicals are both reasonably constant, however, results will be favorable provided dosage is maintained uniformly.

Dosage control requires first an accurate measurement of wastewater flow. This signal must then be linearized (if it is nonlinear) and applied to the chemical feeder or flow controller through a ratio or dosage adjustment. Lime feeders (such as that shown in Figure 48) usually have a numerical but uncalibrated proportioning adjustment. Carbon-dioxide feeders typically have an uncalibrated proportioning adjustment, but a gas flow indicator is available to verify the dosage.

The most positive—but also most expensive—method of chemical feeding uses a flowmeter and controller, as shown in Figure 47. This controller may be set to deliver a fixed flow rate, or may be set in proportion to wastewater flow. The proportioning or ratio setting may even be calibrated in terms of mg/l lime dosage if the lime slurry is of uniform composition. A positive calibration like this is very helpful to operators in making adjustments based on laboratory analyses of either raw wastewater or treated effluent.

Figure 50 shows how to convert an open-channel flow measurement into a linear signal. Rectangular weirs and Parshall flumes produce a head that is related approximately to the 2/3 power of the flow. The differential pressure (dp) transmitter sends the head signal to a square-root extractor, which produces the 1/2 power of head; this signal, when multiplied by head, yields the 3/2 power, which is linear with flow. The exact relationship between head and flow varies from 1.52 to 1.58 power, depending on the width of the throat. Using the 3/2 power results in an error of less than ±1%.

At this writing, pH control represents the most effective control method over phosphorus concentration in a lime-treatment system. Although the relationship between the phosphorus content and pH is not exact, as the scatter in Figure 46 indicates, the pH nonetheless governs its removal. Secondly, it is the only available measurement which indicates that a sufficient quantity of lime has in fact been added

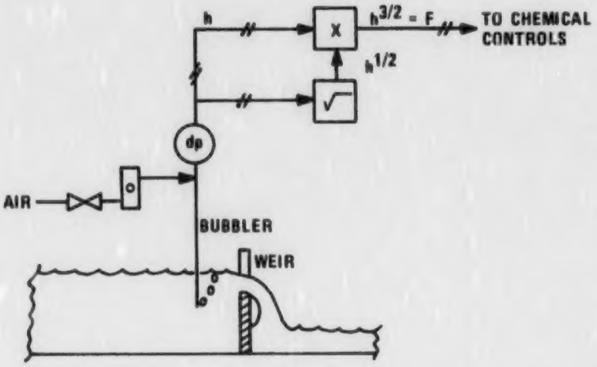


Figure 50. Linearization of head signal from a flume or rectangular weir may be performed with a square-root extractor and multiplier.

to affect precipitation of $Ca_3(PO_4)_2$. Thirdly, pH control can adjust the dosage to compensate for variations in wastewater alkalinity and phosphorus content.

Although pH measurement and control have not found favor in treating water, there were very good reasons for past failures. Failure modes for pH-measuring systems are cited and recommended practices to improve reliability are listed in References 15 and 28. If these recommendations are followed, the pH measurement can be as reliable as any other in the plant.

A particular problem in phosphate removal and recarbonation is the fouling of electrodes due to precipitation. A coating of sludge can build up on the electrode surfaces, which can destabilize control in a few hours. Fortunately, ultrasonic cleaning is quite effective on these deposits.

Figure 51 is a record of the pH and lime-valve position in a laboratory-scale phosphate removal process. Approximately 12 hours after startup, an oscillation began to develop, and expanded until the valve was closing fully each cycle. The reason for the increasing cycle was that the coating insulated the electrodes from the process, creating an increasing delay in response. When an ultrasonic cleaner mounted on the

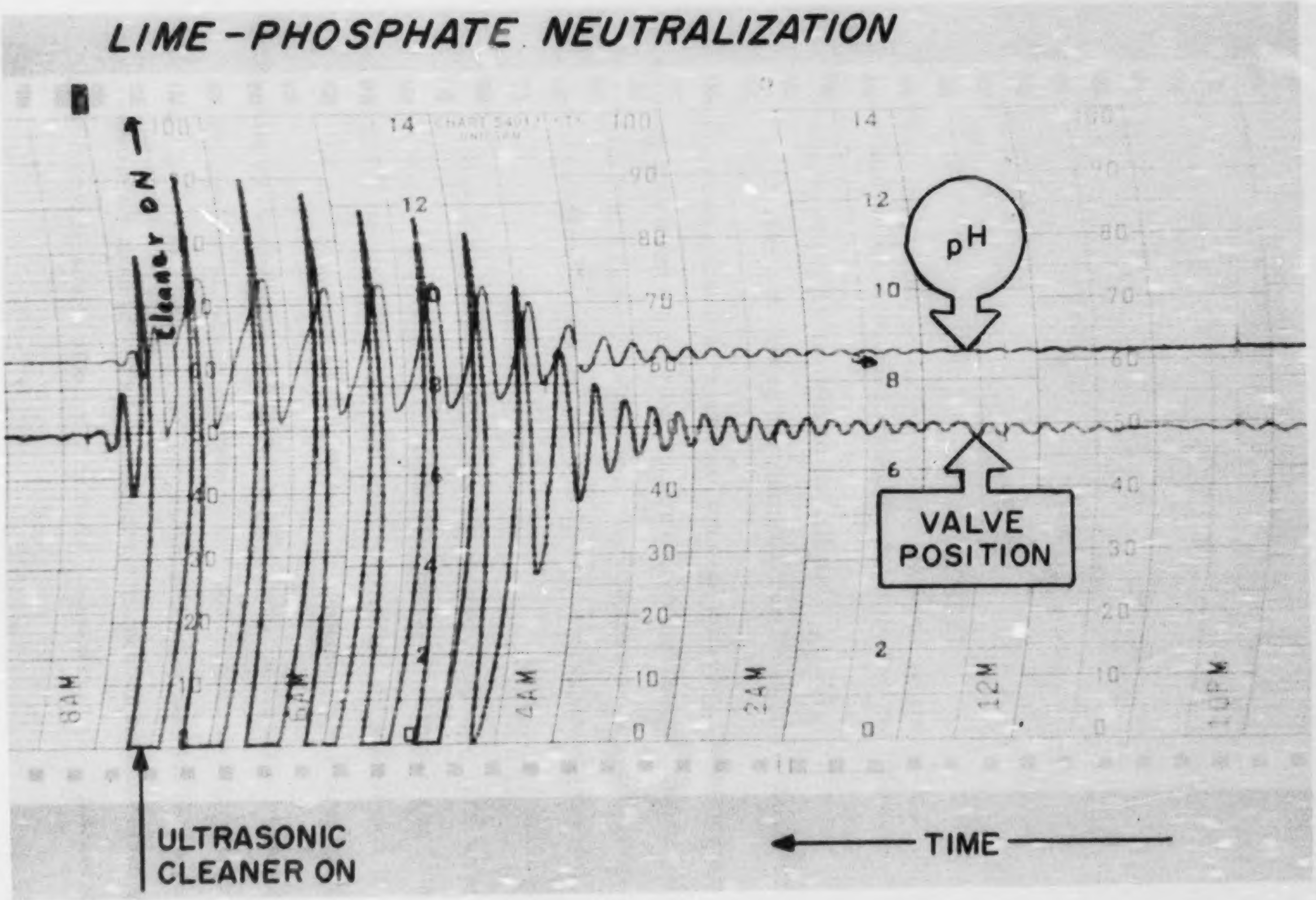


Figure 51. Instability in this record was caused by an accumulation of a coating on electrode surfaces, which was removed by an ultrasonic cleaner.

electrode assembly was energized, the electrode response improved almost immediately and stable control was restored. (Note that time proceeds from right to left in the record.)

In actual practice, the cleaner is continuously energized, so that deposits cannot accumulate. The laboratory test simply demonstrated that the cleaner could also remove accumulated deposits, thereby indicating its effectiveness.

Figure 52 illustrates a pH control system wherein the pH controller sets the lime dosage rather than setting the lime flow directly. This system can be operated in the constant-dosage mode by placing the pH controller in manual. In fact, an indicator connected to the output of the pH controller can be calibrated in any convenient dosage units such as mg/l or lb/million gallons. Then an electrode failure will not result in complete loss of control but simply fall back to the next best system; i.e., constant dosage. (Pneumatic controls are shown, although electronic may be used as well. Conversion from electronic to pneumatic is usually required somewhere in the loop, in that the pH and magnetic flowmeters are electronic while pneumatic valves are customarily used.)

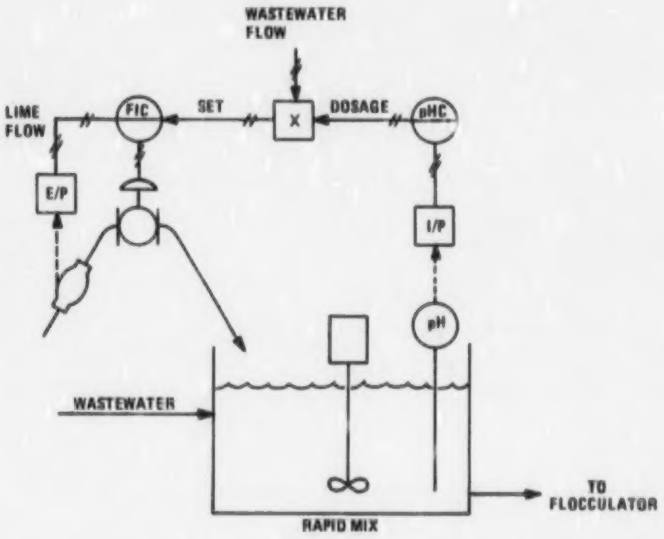


Figure 52. pH control can be combined with dosage control for effective backup.

The titration curve of lime against pure phosphoric acid in Figure 53 shows the range between pH 9 and 10 to be virtually linear. Therefore, a conventional proportional-plus-reset controller may be used in contrast to the nonlinear controller recommended by the manufacturer (28) for neutralization of acids and bases. In two-stage plants (first stage, pH elevated to 11 with lime; second stage, pH reduced with CO_2) where the pH must be elevated to 11, the slope of the curve is lower, facilitating control. However, precise control is more important at pH 11, or considerable overdosage could result.

The titration curve for wastewater would differ markedly from Figure 53 below pH 8, owing to alkalinity absent in the phosphoric acid sample. The same alkalinity would also affect the Ca/P mol ratio in that bicarbonate ions will consume lime as the pH is raised. Considering that wastewaters typically have a higher alkalinity than phosphorus content, this curve is given only as a guide. The alkalinity moderates the curve, tending to make it become more linear.

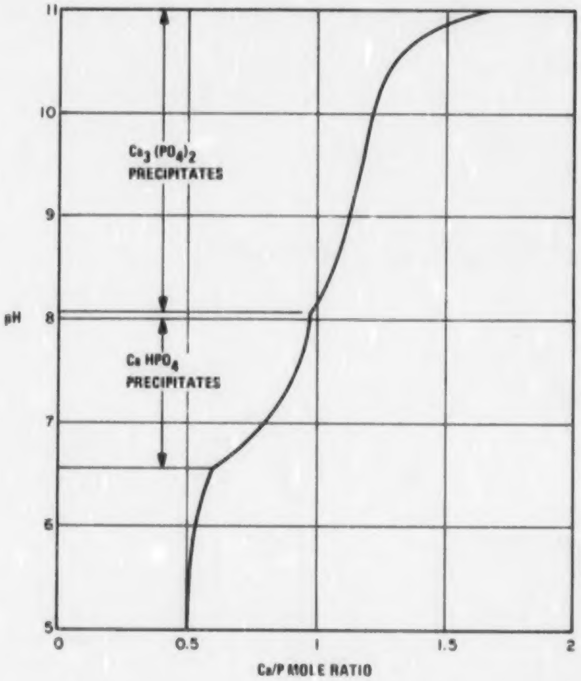


Figure 53. Titration curve of 10^{-3} MH_3PO_4 with $Ca(OH)_2$.

The manufacturer (28) indicates that the residence time between the point of lime addition and the pH measurement should be 15 minutes or more for complete reaction. This limitation applies to neutralization, but for adjustment to pH levels above 9 with lime only two minutes is required (15). Where less than two minutes is available in a rapid mix, pH electrodes should be moved part way into the flocculator to provide the necessary reaction time.

The pH required to produce a given phosphate in the effluent is a function of the calcium-ion level. In systems treating wastewaters of variable hardness and alkalinity, monitoring the effluent calcium-ion level may allow occasional adjustment to the pH setpoint. Since log $\left[PO_4^{3-}\right]$ increases with pH and log $\left[Ca^{2+}\right]$ increases with pCa, an increase of 0.1 pCa should allow a reduction of 0.1 pH while yielding the same total phosphorus. Frequent or automatic readjustment of the pH in this manner could result in lime savings proportional to the variations in calcium content of the wastewater.

Other methods of phosphorus control have also been considered, but, at this writing, an ion-selective electrode for orthophosphate has not been commercially marketed, although there is a demand for such an electrode.

Automatic spectrophotometric chemical analyzers are available for measuring orthophosphate. They may be operated batchwise or continuously, yielding results with a delay of only 5 minutes or so following introduction of a sample. Control could be applied based on this analysis, with the PO_4 controller adjusting the setpoint of the pH controller. However, phosphorus compounds other than PO_4 will escape analysis. Although PO_4 control would be more precise than pH control or pH + pCa control, it still will not provide total P control. The last requires a more difficult and time-consuming analysis, which does not lend itself as readily to control. Experience in treating a given waste may indicate that total phosphorus and orthophosphate are identical or at least correlatable, in which case the simpler orthophosphate analysis is adequate.

An orthophosphate analysis of the wastewater prior to lime addition could be used to set the lime dosage in a feedforward manner. But the lime dosage is primarily determined by wastewater alkalinity and hardness, not orthophosphate concentration. Therefore, unless the prior history of the alkalinity and hardness of the water is taken into consideration, dosage control based on orthophosphate concentration times wastewater flow would be only a slight improvement over dosage control based on flow alone. Both are inferior to pH control.

Recarbonation of treated waters with a pH greater than 10 can reduce alkalinity by precipitation of $CaCO_3$. However, recarbonation must be controlled precisely to achieve minimum alkalinity. The pH corresponding to minimum calcium solubility is approximately 10. It has been reported (25) that a pH of 10.3 gave the most $CaCO_3$ precipitation at the Lake Tahoe tertiary treatment plant. Any individual plant may have its own optimum pH, based on the wastewater characteristics.

Insufficient recarbonation will not remove all possible $CaCO_3$, while excess recarbonation will redissolve some calcium by forming more HCO_3^- ions. Consequently, whatever the optimum pH, it must be controlled precisely because it will yield the fastest settling floc and the least alkaline effluent. Calculations of calcium solubility give concentrations of 2×10^{-4} M at pH 10.5, 1.5×10^{-4} M at pH 10, and 1.8×10^{-4} M at pH 9.5, indicating the significance of precise control.

A conventional proportional-plus-reset controller may be used to control the recarbonation pH by manipulating the CO_2 source, as shown in Figure 54. Both ultrasonic cleaning and automatic temperature compensation of the pH electrodes should be provided for the precipitation at pH 10. However, neither of these features is necessary for recarbonation to a final effluent pH below 8.5.

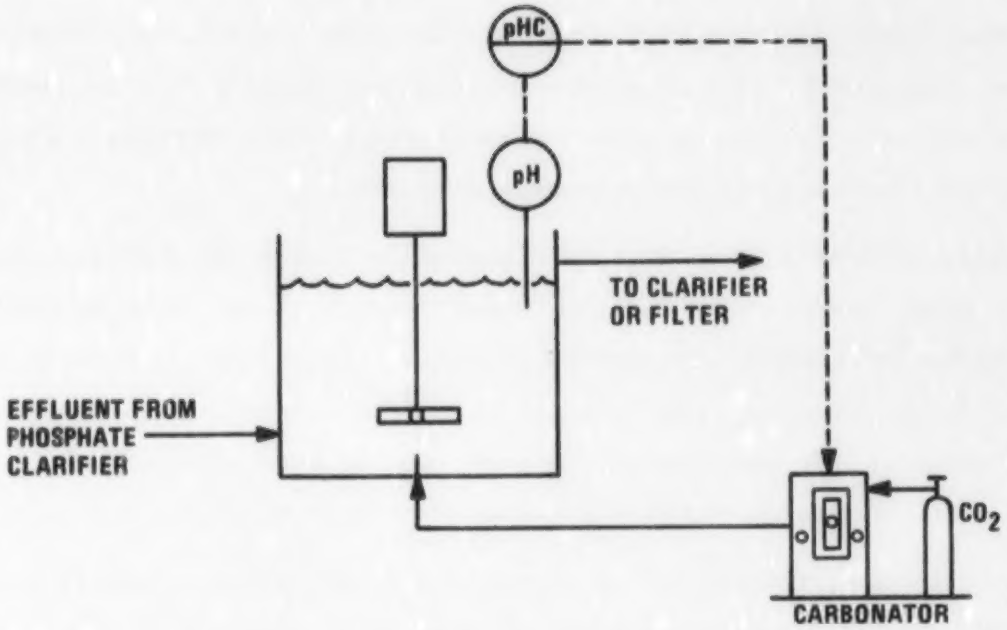

pHC
pH
TO CLARIFIER
OR FILTER
EFFLUENT FROM
PHOSPHATE
CLARIFIER
CO_2
CARBONATOR

Figure 54. pH control applied to recarbonation.

SECTION VII

CONTROL OF WET-WEATHER TREATMENT PROCESSES AND COLLECTION SYSTEMS

Combined sewers make up approximately half of the sewer systems in the United States. Some 1300 U.S. towns and cities are serviced by sewer systems that overflow during rainstorms, when the combined volume of sanitary waste, infiltration, and storm water exceeds the capacity of permanent facilities.

To accommodate these excess flows and to prevent the flooding and polluting of sewered areas, most combined sewers direct the excess flows through overflow structures into neighboring watercourses. The overflow structures route the small dry-weather flows through intercepting sewers to treatment facilities. (Automatic mechanical regulating gates have been used widely to make improved separations between dry-weather and storm flows, with varying degrees of success (29).)

Although they may only make up a small fraction of the total annual volume of sewage handled, overflows are a major source of water pollution. Whether from sewers or treatment plants, overflows (bypasses) usually account for a disproportionately greater quantity of the biologically active and toxic materials being delivered to nearby streams (30). Since the often-thought-of direct solution to combined sewer overflow pollution, that is, separation, would be extremely costly and require several decades to achieve, sanitary engineers have proposed alternative measures that would be effective sooner at only 40% to 70% of the separation cost (31). These methods (based on various combinations of storage and treatment processes, followed by disinfection) are discussed at length in this section. Another advantage of the alternative measures over sewer separation is that all flows into the receiving body (or bodies) of water will then be treated, whereas the effluent from a separate storm sewer usually is not treated (32).

Tests have shown that some form of primary treatment is usually sufficient to reduce organic and solids pollution levels to allowable limits. Since it rains only 4 to 15% of the time, storm-water facilities are designed with low capital and high operating costs (just the reverse of the criteria used for most treatment plants).

The topography and physical layout and size of the collecting watershed, the sewer system, and the availability of land at the proposed site all enter into the selection of storage and treatment methods; e.g., off-line storage, routing and storage in the sewer, handling in the satellite treatment plants, or centralized treatment. Because all treatment plants provide some degree of storage, the relative storage size and treatment plant rate desired depend largely on the nature of the expected overflow.

The influent to an overflow treatment plant results from a transient storm-water flood that is far in excess of normal variations in dry-weather sewage flow. Overflows in both combined and separated sewers are characterized by a large and rapid rise in sewage flow and strength (33). It is this sudden mass of highly contaminated water that must be prevented from polluting local watercourses.

As a specific treatment approach, microstraining has been shown to be effective in reducing solids and BOD from overflows (34, 35). Other methods for handling overflows include in-line storage, retention basins, and centralized overflow treatment and are discussed in References 29, 34, and 36 through 39.

The concept of automating storm and combined sewer systems will be explained below by considering three hypothetical cases. Each case will consider a different size catchment area. The treatment systems discussed in each case are hypothetical solutions to the problem and other treatment systems may also be applicable to handle the storm waters within a catchment area. The three catchment areas selected for this study are 1) small (150 acres), 2) medium (3000 acres) and large (100,000 acres).

COMBINED SEWER OVERFLOW TREATMENT PLANT SMALL (HYPOTHETICAL 150 ACRE) CATCHMENT AREA

In the small catchment area microstraining offers a practical method to achieve the physical treatment of combined sewer overflows. The hypothetical treatment facility that makes use of such a method is generally relatively small and self-contained, and usually is located at an overflow point that is relatively remote from a central treatment facility.

In this section the combination of storage and microstraining as a suitable alternate method for treating combined sewer overflows will be discussed. The basic instrumentation for such plants is reasonable and well established, and will operate effectively if properly installed and maintained.

The process objectives are as follows:

- To eliminate pollution (caused by combined sewer overflows) by collection and sufficient treatment to allow a discharge into the receiving bodies without causing their degradation
- To handle the specified design overflow with minimum capital and operating costs
- To provide a safe, practical, and efficient facility
- To consider a treatment plant that utilizes microstrainers and is suitable for a relatively small catchment area.

Background

The overflow event is marked by a relatively sudden variation in flow rate and in the concentration of pollutants. From a hydraulic viewpoint alone, overflows can be several hundred times the magnitude of the normal dry-weather flowrate (40). Pollutants classifiable as suspended solids and BOD or COD are usually strongest at the start of the overflows and may contain difficult contaminants such as grease or salinity, but this depends entirely on local conditions. The overflow itself is initiated by either rain or thaw; consequently, typical rain and runoff patterns must be examined as part of the overall problem.

The quantity of precipitation and subsequent events in the sewer system must be measured to a finer time scale than is usually customary in computing hydrographs for hydraulic engineering studies (41). These studies have to be reasonably accurate for an entire seasonal cycle as a minimum. In any case, a statistically valid picture of the expected overflow must be produced if the facility is to be designed to cope properly with the event.

Because the overflow reaching the treatment facility is always a highly dynamic occurrence an initial storage volume is desirable. Initial storage also provides the means for grit and other solids sedimentation that can reduce the burden on subsequent treatment facilities. The solids accumulation can be removed by the most convenient and economic method after the overflow has subsided. If treatment by a downstream continuously operating plant is preferred to treatment by the overflow facility, the entire content in storage can be returned to the downstream plant as soon as the overflow has ceased.

The usefulness of a microstrainer in overflow treatment for small catchment areas has been described in U.S. EPA publications (34, 35) and is summarized below. The microstrainer operated at flow rates of 35 to 45 gallons/minute/square foot with differentials of 24 inches of water. At these rates, the suspended solids in the combined sewer overflow were reduced from an influent range of 50 to 700 mg/l to an effluent range of 50 to 40 mg/l and below. At the higher influent levels of suspended solids, the removal performance was enhanced, yielding effluent concentrations of approximately 10 mg/l. The conventionally used percentage removal performance criteria are not valid for microstraining the combined sewer overflow. The volatile suspended solids reduction paralleled the reduction of total suspended solids.

The highest concentration of suspended solids frequently occurs when the overflow rate is highest. The concurrence of a high suspended solids concentration and a high overflow rate results in a very high potential of contaminant loading per unit time into the reciving stream. The microstrainer was unusual in that it removed a much greater percentage of the suspended solids when their concentration was higher. A reasonably sized microstrainer, then, can limit the pounds of suspended solids per unit time entering the stream from a combined sewer outfall.

The organic matter in the combined sewer overflow was highly variable—ranging from 10 to 2000 mg/l as BOD_5, and 20 to 4000 mg/l as COD. The microstrainer reduced organic matter by some 25% to 40%.

Satellite Overflow Treatment Plant

This hypothetical proposed facility is considered to be semimanned in that it is sufficiently automatic to detect and handle any overflow but requires routine and conscientious maintenance, as well as manual cleanup and replenishment after an overflow. As outlined in Figure 55, the facility consists of the following:

- An inlet structure
- The storage and surge reduction volume (if found to be necessary)
- Treatment facilities (microscreens)
- A disinfection system with contact provisions
- A disposal system and miscellaneous pumps, equipment, and controls
- The station itself (consisting of buildings, etc.).

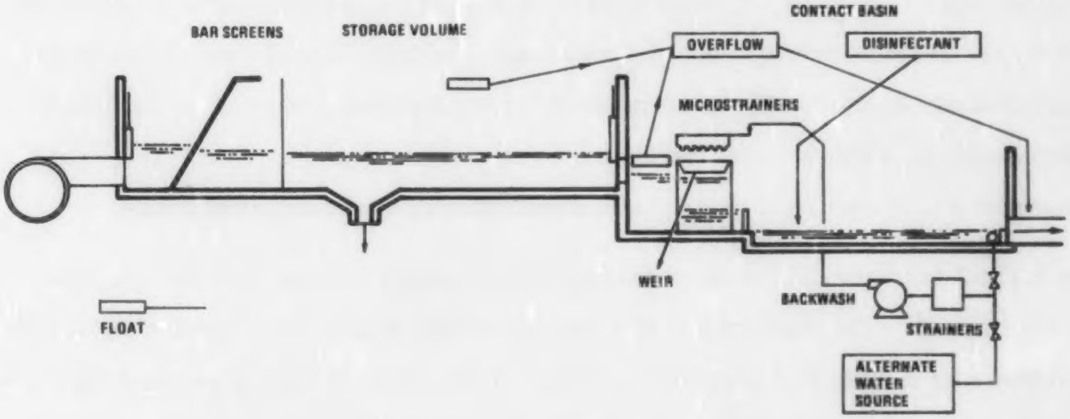

CONTACT BASIN
BAR SCREENS
STORAGE VOLUME
OVERFLOW
DISINFECTANT
MICROSTRAINERS
WEIR
BACKWASH
STRAINERS
FLOAT
ALTERNATE WATER SOURCE

Figure 55. Satellite treatment plant.

The inlet structure, which accepts the overflow, contains gates for station isolation and will also include pumps if necessary. Coarse racks and other devices to handle large objects and protect the downstream machinery will be located either here or in the intermediate structure. A sensing device to detect the overflow start may also be located in the inlet structure.

The storage and surge reduction volume accepts the peak flow rate of the overflow as effluent is withdrawn through the microstrainers at a maximum rate. A storage volume may not be necessary in all cases, but if storage is provided some sedimentation will occur. The storage volume is designed to be kept nearly empty between overflows and has a residual volume that will retain sludge and isolate it from the microstrainers.

The residual volume, tank washings, etc., are usually pumped back to the interceptor or main treatment plant as sludge. (Sediment might be removed by earth-moving machinery at less cost, in some cases.) The storage tank also contains an emergency overflow to prevent flooding the station. Scum, grease, and other clogging materials require special handling to prevent fouling the microscreens, but the cure would have to vary from situation to situation.

The intermediate structure separates the storage volume from the microscreens. This structure contains gates to moderate the flow from the storage volume to the microstrainers and also allows isolation of one or more microstrainers. It may contain bar screens, pumps, or chemical dosage systems, including flowmetering.

The disinfection system is required to treat the sanitary component of the overflow for health considerations. The disinfectant addition rate must be maintained proportional to the effluent rate, and the treatment ratio is readjusted by a suitable residual analyzer system. The contact tank is sized to ensure sufficient residence time for disinfection under maximum flow conditions, and also provides a source for reasonably clean water for screen washing and station cleanup. Standby water is required if the contact basin runs dry or becomes contaminated by solids, due to an overflow or screen failure. If the standby water source consists of city water, a backflow preventive device is also required.

Sludge disposal pumps and auxiliaries are required to wash down a facility during and after an overflow and to assist in manual operation. Controls will be required, too, for the throughput rate (via level control), microstrainer, operation, disinfection, and alarms. Recording and communications systems are also necessary.

The microscreen section, containing two or more machines, has its own control system, which varies the drum speed and backwash flow in response to the head loss. A source of spray water and a drain for drum washings are both required. A high differential alarm detects a hydraulic overload condition.

Microstrainers have a limited maximum flow capacity (typically between 10 and 40 gallons/minute/square foot of the submerged strainer area) when operated at high pressure differentials (42). So long as that differential is not exceeded, they are self-regulating. The water level on the outer or downstream side of a microstrainer is maintained at about one-fourth the diameter of the drum in order to maximize the differential and to ensure that the material on the screen does not dry out. The upstream level is kept nearly up to the backwash overflow level to maximize the effective filter area and head across the screens (42).

When the input level to the microstrainers is kept at a fixed value, some means must be provided to control the flow from the storage volume as the storage level rises to provide more capacity for the overflow. Where head is available, automatically positioned gates are suitable; otherwise, pumps could be used in the usual fashion for level control. The microstrainer controls that are usually furnished (to control the drum speed and backwash rate in response to the differential) are less important when the input level is controlled, but they do provide the means for setting lower drum speeds manually if desired for improved solids retention or better backwashing.

It may be desirable to improve the solids removal capability through pretreatment by adding a coagulant but such a procedure, though feasible, would depend on field trials. In any case, a suitable and continuing sampling and testing program is required to control performance, both in the design phases and after the plant is built and running.

Disinfection systems based on applying disinfectant in ratio with effluent flow, with a residual analyzing trim, are well established in practice and need no discussion here. The rangeability of such a system is limited, however, and the effluent flow rate variation should not be greater than about 10:1, using a single disinfectant feed system. (This is based on a chlorinator rangeability of 20:1 and a dosage rate variation of 2:1.)

From an operational standpoint, a station should be sized so that the treatment portion runs continuously and all overflows are contained in an intermediate storage. This would use the treatment system most economically. Sizing the station by using extensive storage to handle all storms is usually impractical; the station should be designed to treat overflows as efficiently as possible within a maximum practical storage capacity. The treatment efficiency will be maximum if, as soon as the storage volume has accumulated to a certain minimum, the treatment system is made to run at its maximum rate and continues to run at that rate until the overflow ceases or the volume in storage is at or near its minimum.

Facility Sizing

The 150-acre drainage area is an indefinite measure for overflow treatment plant sizing. The overflow facility cannot be sized for fixed flow rates but, in any area, will depend upon the:

- Typical local rainfall—its duration, intensity, and frequency data (more detailed in space and time (43)
- Dynamic analysis of the catchment area
- Dynamic analysis of the sewer system between the catchment area and the proposed plant
- Dynamic interaction between the proposed facility and the existing system.

A computer-type program is the logical method for performing sizing calculations, but only after data has been collected in the area.

A design runoff rate of 1 cubic foot/sec-acre (44), produced by a 50% collection (runoff coefficient = 0.5) from a 2 inch/hour continuous rainfall, is equivalent to a continuous runoff rate of approximately 100 mgd from a 150-acre catchment area. The following discussion is patterned on an estimate of a suitably sized plant for a noncontinuous situation.

A 2 inch/hour storm of 1-hour duration produces 54,300 gallons per acre. If the runoff coefficient is 0.5 from the 150-acre watershed, then the total runoff is about 4 mg. For the hypothetical example, assume an overflow where, due to the rainfall distribution (in space and time) and the sewer characteristics, about 2.8 mg reach the satellite treatment plant, with a peak influent rate of approximately 46 mgd (Figure 56).

Further assume a 2 mg storage volume, followed by a 25 mgd capacity treatment system. As noted in Reference (36), there is a tradeoff between storage volume and treatment rate or in other words the plant size and the storage volume are interrelated. In the reference cited, the ratio of volume to rate ranges from 0 to 0.286 day; that is, the plant will be operating on an average basis that amount of time. In this example, the time is chosen as 0.08 day.

Treatment of the accumulated overflow is arbitrarily assumed to start 30 minutes after the overflow begins filling the storage volume, when the stored level has reached 10% of capacity. Without level control (which allows the water to flow to the microstrainers through a fixed restriction at a rate proportional to the level), 65% to 70% of the storage capacity is used and treatment lasts about 6 hours. With level control (which allows the water to flow to treatment at the maximum rate of 25 mgd), less than 40% of storage is used and treatment lasts only 2.6 hours. For these reasons, the volume and treatment rate selected are a reasonable choice for an overflow of such dimensions.

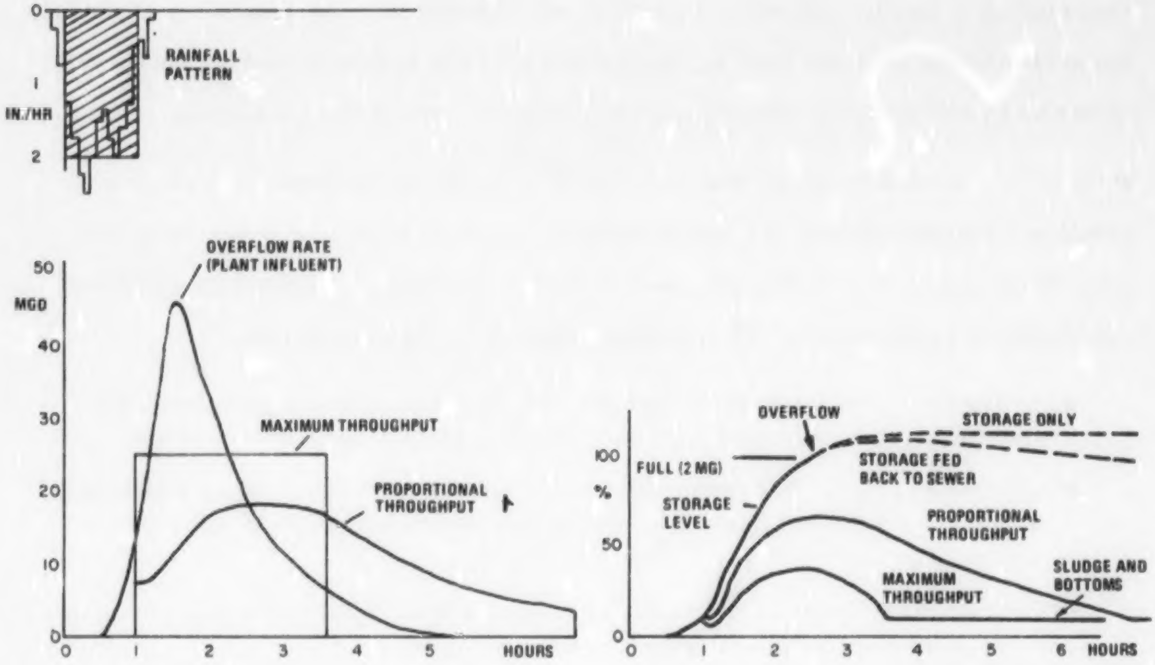


Figure 56. Satellite plant performance.

The use of level control in this hypothetical case greatly increases the potential capacity of the station. Precise capacity is hard to estimate without more factual information about the overflow since an actual situation is almost certain to be non-linear in several ways, but a suitable combination of representative field data and a dynamic analysis should be sufficient. (An analog computer or its digital equivalent should be considered for this type of calculation.)

Control Objective

Overflows of excessively high rate and duration can exceed the storage volume and the throughput rate of the treatment facility. In these cases the excess flows should be allowed to overflow in the least objectionable manner before the facility itself is flooded.

Disinfection is easily controlled through standard methods. The feasibility of closed-looped control of chlorine residual has previously been demonstrated for continuous plants and, with some modification, is applicable to intermittent automatic service.

Because of the remote nature and multiplicity of installations implicit in the use of satellite treatment plants, it is most desirable from an economic viewpoint to automate these plants so that they can operate largely unattended, with no loss in either reliability or performance. The principal objectives, then, are that:

- Instruments and instrument systems shall be chosen as components of the instrument arrangements proven to be workable in wastewater systems.
- Control shall be either pneumatic or electronic, as dictated by system capability. The high reliability and low cost of pneumatic systems must not be ignored.
- Instrument systems selected must be capable of a full and accurate performance upon energization after long shutdown periods.
- Equipment shall be selected for reliability, minimum maintenance, and performance—in that order. Accuracy is considered less important than repeatability and reliability and need be no more than necessary for system functioning as a whole.

From the viewpoint of facility functioning, it is desirable to monitor the throughput rate and the concentrations of the constituents of interest as a function of time. For the purpose of operating the station (either automatically or manually), the following measurements are desirable:

- Influent channel level to initiate the station operation
- Storage level to warn of a possible overflow and to pace the throughput rate (if desired); also an alternate for station startup
- Individual strainer differential to control the drum speed and backwash rate and to warn of a possible overflow
- Influent or effluent flow rate to compute the strainer performance, to control the disinfection rate, and to maintain a historical record
- Disinfectant process concentration measurements in the effluent to ensure compliance with standards

- Organic strength to ensure compliance with standards
- Solids concentration to ensure compliance with standards and to indicate the strainer performance
- Salinity concentration or a conductivity measurement
- Facility internal status checks, illegal entry by unauthorized persons, fire and flood alarms.

For control, the mechanical and hydraulic measurements, chlorine residual, and salinity are the only ones practical at this time (45). Automatic turbidimeters or nephelometers for solids determination are in a marginal performance status, although they have had some success in estimating solids content. Automatic sampling is available for accumulating samples for manual laboratory analyses (especially when such automatic sampling is based on sound principles), but automatic instruments for suspended solids or organic content (especially those suited for unattended locations) are not yet commercially available.

Automatic controls are necessary for the rotary strainers, the disinfection system, the level control to the strainers, and station management if the satellite station (whether manned or not) is to be able to handle overflows as they occur.

Microscreens are not installed singly, since doing this might put the facility out of service when a screen was down for maintenance. It may prove desirable to limit by automation the machines in service so that only a minimum number will operate under minimum flow.

The disinfection system may not be suited for operation at very low flowrates. Consequently, some means for maintaining the facility effluent rate above a certain minimum is desirable; it is for this reason that level control between the initial storage and the microscreens is similarly desirable. For the same overflow event, level control (or pumps) between the storage and screens could allow some 20% reduction in the storage volume. The exact benefit would depend on the process parameters, especially the nature of the overflow incident.

A signal for treatment initiation is advisable to start facility operations. Normally the facility operates either in a standby or an active mode. Manual maintenance and post-event cleanup are assumed. The signal for treatment initiation could be sent to a remote site for automatic notification.

Plant Operation

In a normal operation between overflows, the facility remains in a standby condition. The inlet gate is open, the storage tank is empty, all motors are shut off, and the screening and disinfection systems are in condition to start. Wherever necessary (as in an instrument that depends on electrochemical probes), a small flow of clean water may be maintained through the instrument to keep it in readiness.

The facility itself is secured against intruders, fire, freezing, etc., with all necessary alarms and detectors and all instruments in a standby condition. Routine maintenance is performed as necessary, and possibly various systems are exercised after extended dry periods.

At the start of an overflow, the level in the storage tanks starts to rise, the recording instruments start, and an alarm is initiated at a remote control center. After the level has risen sufficiently to start flowing to the screens, a screen or screens are started at minimum speed, and backwashing at the top of the screen begins at a minimum flow. As the flow continues through the screens, the effluent flow is sensed and the disinfectant sampling systems start. As the overflow continues, the level at the input to the screens rises, the flow to the screens increases, the head loss across the screens increases, the drums turn faster, and the backwash flows increase. The increased effluent flow increases the disinfection rate, although it is modified by the chlorine residual control system. The sampling system collects effluent, either at a fixed rate or in ratio to the main effluent rate. As the level in the storage volume continues to rise, the level at the input to the screens reaches its operating maximum and is prevented from rising further by the level control system and the automatic control gates. The screens then operate at constant influx until the overflow begins to end. In this way the storage volume accumulates the overflow up to its maximum capacity, without causing an overflow at the input to the screens.

At the end of the overflow as the level in the storage begins to drop again, the screens and disinfectant systems slow down after the level has dropped below the input level to the microscreens. Eventually the screens stop and the station returns to a standby position until it can be cleaned up, replenished, and made ready for the next overflow event.

The facility will operate reliably in this way only if maintenance and servicing are correctly carried out to the degree determined by the original design, as well as by experience.

Instrumentation

Instrumentation for a satellite facility must be simple to operate and maintain, highly reliable (under existing conditions of service and maintenance available), failsafe, quickly repairable or replaceable, and inexpensive. All criteria can be met by using the conventional electronic or pneumatic instrumentation common to the process industries.

Instrument interconnections must be of high quality and installed in accordance with good practice. Wherever explosive hazards are absent, NEMA 4 housings, sealed conduits, and noncorrodible materials should be used for electrical systems. Wherever conditions indicate that electric power is potentially hazardous, pneumatic or intrinsically safe systems are strongly recommended. Pneumatic systems must be designed to resist both corrosion and leakage.

The instrument panel must be properly designed for use and maintenance, with all connections well marked and secure but easily accessible. Power supplies for electronic systems must have a conservative safety factor and be self-protective.

Instrument air systems must produce clean and dry air. In any case, the system should have a guarantee of at least 18 months between failures, assuming that a specified and controlled maintenance program is set up and followed.

Due to long periods of inaction, the recorder will be shut off when the station is idle. The chart could be automatically marked with a date/time stamp when the overflow begins, or an intermittent strip chart could be used. The recorder has a uniform percentage marking for all three inputs, and the time scale is measured from the starting time for that occurrence.

The level measuring systems for the storage volume and microstrainer input (LT-2 and LIC-3) are analog devices and should be of the bubbler type, with flow control relays, manual purge valves, and some means for automatically shutting them off when the facility is idle. The inlet level detector (LS-1) is properly an inverted bell or diaphragm box, capable of operation without an air supply. Conductive or capacitive probes must be carefully applied since intermittent usage, dirt, and moisture have caused many such installations to fail in the past. Redundancy is advisable here.

The level control system (loop 3 of Figure 57) senses the level at the common inlet basin for the microstrainers and raises or lowers the motorized sluice gate as necessary to maintain that level. A position indicator (Z1-3) shows the gate position on the instrument panel from which an operator can control it. Also shown in Figure 57 is an interlock (LS-2B) that would keep the sluice gate closed whenever the level in the storage volume was too low, and could also start the microstrainers. LIC-3 is a narrowband proportional controller. The proportional band, sluice gate size and speed, and controlled basin volume should all be sized by a control engineer.

The controlled gate driven by LIC-3 must have a feedback slidewire suitable for wet conditions; in addition, it must be designed for frequent actuation and for long periods of idleness. It is assumed that the design control engineer and the commissioning instrument engineer have ensured that the loop controlling the gate (or whatever device is used) is not too sensitive and will not immediately wear itself out.

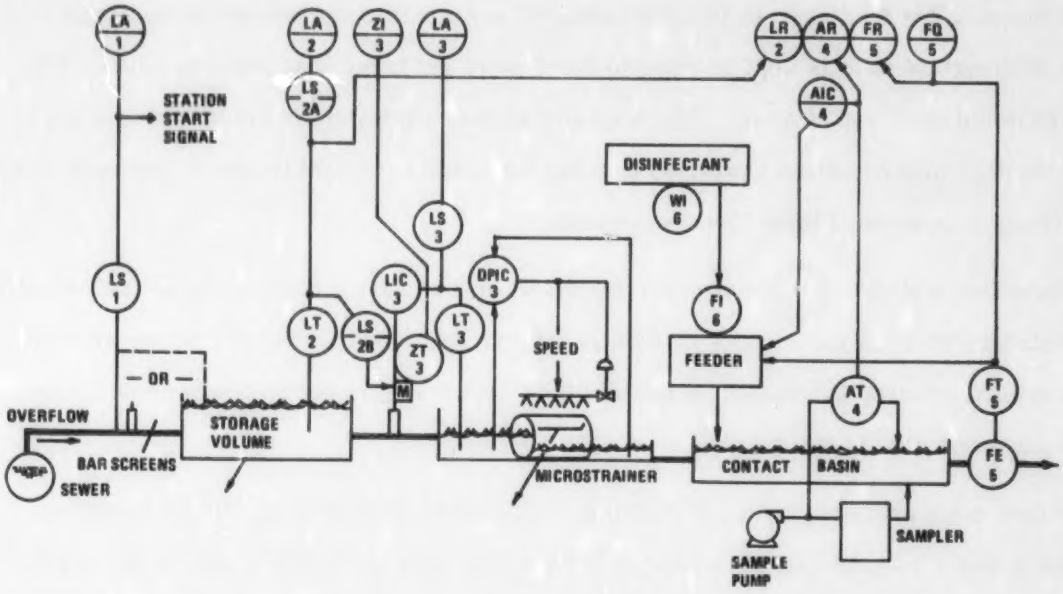

LA 1
STATION START SIGNAL
LS 1
OR
OVERFLOW
BAR SCREENS
SEWER
STORAGE VOLUME
LA 2
LS 2A
LT 2
LS 2B
ZI 3
LA 3
LS 3
LIC 3
ZT 3
M
LT 3
DPIC 3
SPEED
MICROSTRAINER
DISINFECTANT
WI 6
FI 6
FEEDER
LR 2
AR 4
FR 5
FQ 5
AIC 4
AT 4
FT 5
FE 5
CONTACT BASIN
SAMPLE PUMP
SAMPLER

Figure 57. Satellite treatment plant instrumentation.

The microstrainer controls (dPIC-3) are sized and furnished by the machine supplier. The controller is either proportional or proportional-reset with anti-windup. The backwash valve must be capable of handling dirty water and producing a backwash flow that varies linearly with the controller output.

Residual chlorine analyzers are reliable when correctly installed and maintained, but they must be handled properly. The sample stream must be representative, fresh, and reliable. The relatively large sample pump (typically 1 to 2 hp) is a good source, but the sample must still be reliable after passing through an adequate strainer. (The strainers customarily supplied with these analyzers are designed for clean water service and consequently are apt to prove inadequate for this service.) The sampling point must be properly chosen (46) and some provision must be made to prevent the sensing cell or probes from degrading while the facility is idle.

The disinfectant feed system is a conventional chlorinator system or its equivalent. The disinfectant rate is kept in ratio to the flow (FT-5), but that ratio is adjusted by the residual analysis (AIC-4). The feed device has a minimum rate at which it adds disinfectant to the contact basin, so that the feed device should be set to run only when a certain minimum effluent flow is present.

Disinfectant addition control is available as a package that includes a scale (WI-6) and flowmeter (FI-6) in the case of chlorine systems and needs no further discussion here. The typical chlorinator can be modified either at the factory or in the field to accept an analyzer trim signal, as well as the flow signal.

For flow measurement, a Parshall flume with a stilling well is the normal choice, since it has a rangeability in excess of 20:1 and is simple, rugged, and inexpensive.

Instrument Costs

This hypothetical facility provides a 2 mg storage and 25 mgd treatment rate. Note that this size facility, starting empty and beginning the treatment when 0.2 mg has been accumulated, accommodates a constant flow of approximately 50 mgd before overflowing at the end of 1 hour and is therefore comparable to a 50 mgd plant (where such level control is not used).

The associated capital cost for installed instrumentation is approximately $40,000 (excluding the connections to remote facilities, microscreens, gates, and chlorinators) and will vary only slightly with the facility's size until growth is accomplished by installing duplicated treatment units. (The instrumentation costs related to microstrainers are included in the cost of the microstrainers.) The annual maintenance cost for satellite systems is not linearly related to size, and typical maintenance costs are about $30,000 per year for a district comprising up to six satellite facilities. Wherever more than four satellite facilities exist in one locality, communication and computerization should enter the picture, and highly specialized technicians (whether on the payroll or contracted) may be required.

OFFLINE STORAGE AND CHLORINATION TREATMENT PLANT (3000 ACRE DRAINAGE AREA)

This hypothetical facility (outlined in Figure 58) consists of the following elements:

- Inlet structure
- Pumping station
- Storage and short-time sedimentation units
- Disinfection system with contact provisions
- Pumpback and sludge cleaning facilities
- Miscellaneous pumps, controls, and equipment
- Buildings necessary to house the facility.

In this concept excess stormflow is captured in a holding basin where gross solids settle out. The liquid is chlorinated before discharge to effect microorganism control.

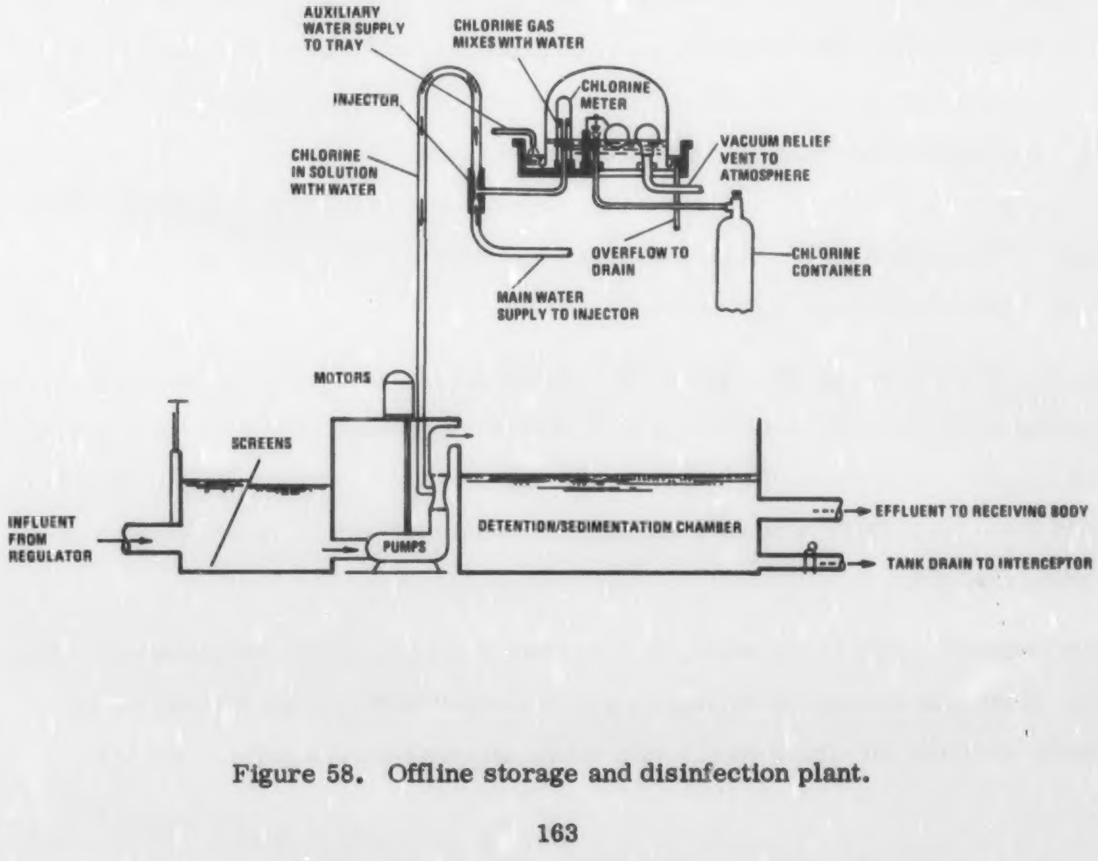

AUXILIARY WATER SUPPLY TO TRAY
CHLORINE GAS MIXES WITH WATER
INJECTOR
CHLORINE METER
CHLORINE IN SOLUTION WITH WATER
VACUUM RELIEF VENT TO ATMOSPHERE
OVERFLOW TO DRAIN
CHLORINE CONTAINER
MAIN WATER SUPPLY TO INJECTOR
MOTORS
SCREENS
INFLUENT FROM REGULATOR
PUMPS
DETENTION/SEDIMENTATION CHAMBER
EFFLUENT TO RECEIVING BODY
TANK DRAIN TO INTERCEPTOR

Figure 58. Offline storage and disinfection plant.

This station is considered to be semimanned; during nonoperating periods the day shift is covered by a custodian/maintenance man. The initiation of the station's operation is automatic, with necessary personnel being summoned in order to be present throughout any storm event. The actual control of the station, too, is automatic, with personnel watching for equipment failures and performing station cleanup upon cessation of the storm event.

This type of treatment facility is structured for a medium-sized drainage area that has inherent regulation due to in-system storage. The satellite facility must be able to respond to rapid changes in flowrate due to the small area served and a lack of inherent system storage. Depending on the nature and size of the area served, it is possible for a very short storm to be over before the installation is activated. It is also quite possible for a small storm not to overflow the plant if the storm's volume is less than the plant's storage capacity.

The inlet structure accepts the overflow from the combined system, probably through a regulator station. The structure contains gates, screens, and a pumping station. The pumping station delivers the storm-water overflow into a series of short-time detention/sedimentation tanks. As the combined sewage is pumped from the inlet structure to the treatment tank chambers, chlorine or other disinfecting agents are added. After short-term sedimentation and disinfection, the plant effluent is discharged into the adjacent watercourse.

The prevailing process philosophy is that storm-water overflows often carry a greater pollutional load than the community's raw sewage. Potential pollutional loading is of such magnitude that short-time detention for the removal of gross solids, followed by disinfection, provides a dramatic reduction in pollutional loading on adjacent watercourses (37, 47).

Upon cessation of the storm event, any overflow remaining in the sedimentation chambers, along with sludge, screenings, etc., is pumped back into the interceptor for ultimate delivery into the central treatment facility with the exception of the few facilities that have on-site sludge handling capability. The sedimentation chambers are not normally equipped with sludge scrapers. The station cleanup and pumpback at the end of a storm event are usually performed by the station personnel who were present during the storm. Some plants such as Spring Creek and Jamaica Bay, N.Y., have automatic cleanup capability.

Following a storm event, the facility is pumped dry, cleaned up, and generally restored to standby status. Because the facility's operation is essentially batch in nature, continuous scrapers are not necessary in the sedimentation basins. These basins could be cleaned in any number of ways, including hand labor and hydraulic flushing.

The disinfection system is designed to maintain a desired residual at the end of the contact period by applying the disinfecting agent in proportion to the flow and by adjusting the dosage through use of an automatic residual chlorine analyzer control system. Chlorine dosages range from over 4 mg/l at low flows to 2 mg/l at high flows; the objective here is to provide a 1 mg/l residual as the treated effluent leaves the facility.

Facility Sizing

Sizing the facility usually is based on the premise that the worst conditions will be encountered whenever the duration of the storm equals the time of the concentration. Whenever rain falls in a given area, it takes time for the water to flow from the periphery of the catchment area to the control point. (The control point is defined as that point which all water from the catchment area must pass.) This total time is known as the "time of concentration." If a storm ends before the flow from the farthest point has reached the control point, the area provides temporary storage and the maximum runoff rate at the control point is less than if the storm had persisted.

It is generally recognized that, the shorter the duration of a storm, the greater will be the expected average intensity. Consequently, the most critical storm is one whose duration equals the time of concentration. A storm of longer duration has a lower intensity, while one of shorter duration will never reach its peak (as far as control works are concerned) before the storm is over. As a result the plant design is a function of the topography and a coefficient of runoff as well as of rainfall. A small hilly area that is heavily built up may have a runoff equal to a much larger area of moderate topography and building density. The result of such unpredictable variables is that facility size depends only partially on the drainage area. The design of such facilities must be structured around the rate of flow rather than the total flow; in brief, each facility presents a unique design problem. The design of the storage capacity necessary for a facility of this type depends on the minimum settling time desired, as well as the maximum design flow.

Control Objectives

The functions of instrumentation contained in this class of facility are as follows:

- To provide control of treatment facilities, chemical feed pumps, and proper sequencing of the sedimentation chambers
- To monitor the station and equipment status
- To indicate and record important parameters for facility and process evaluations.

In a broad sense, the functions denoted above are true of all combined sewer overflow facilities. However, in view of the specific instance at hand and the present state of the art, station records and monitoring functions do provide important aids to evaluating such facilities.

The problems involved in achieving these objectives are many, principally because this type of installation operates discontinuously. This in turn causes a number of difficulties not typical of conventional treatment facilities, such as:

- The equipment stands idle and inoperative between storm overflow events.
- The sampling systems, chemical feed systems, etc., require careful and continuous maintenance if they are to operate when needed.

Since the facility is capable of only partial removal of floating and settleable solids, over-design events deteriorate the effluent inversely as the peak rate during the peak time. Because the process is batch in nature and requires fairly complex sequencing controls as well as chlorination control, process control is necessary.

The manipulation variables are station flow and chlorine flow. Station flow is a manipulation variable in that the flow can be diverted around the station, or diverted to parallel tanks, or both.

Plant Operation

Normal operation between storm events or overflows requires that the facility be maintained in a standby condition. This means that the inlet gates are open, the fine and coarse screens are not running, all motors are off, and the disinfection system is shut down. The residual chlorine analyzer is flushed with potable water to keep the electrode system in an operable condition, and the normal lighting and ventilation circuits are kept energized.

The facility is secured against intruders, vandals, and fire by means of the necessary security and alarms. The facility size and complexity necessitates routine daily custodial service and maintenance procedures. Routine facility exercises are generally not essential because of the frequency of operation and the presence of operators. However, routine maintenance is essential for all mechanical equipment, and most especially for the instrumentation.

At the start of any overflow event, the system overflows through the inlet gates into the screen chamber, where the rising level initiates the screen drives. The overflow continues into the wet well of the pumping station, where the rising level starts the first pump at a low speed. As the level continues to rise, the speed increases over the control band. A continuing rise in the level starts the second pump at a low speed, etc. Upon decreasing level, the pumps decrease in speed and finally drop out in reverse order.

Each pump is equipped with a flowmeter that in turn operates a chlorine feeder in proportion to the flow. In this way, the potential operating difficulties associated with summing systems and sequential feeder operation are eliminated. A flow summing system requires a flowmeter and transmitter for each measured flow. The signals from each flow are fed into a battery of summing relays or summing amplifiers. The signal representing the total is sent to the readout and feed equipment. However, the signal representing each flow has an elevated zero (e.g., 3 lbs, 4 ma dc, and 10 ma dc), which requires that an appropriate zero signal be supplied to the summing relays or amplifiers for each measured flow that is zero. Because the most probable point of error for any instrument is zero, this signal is usually provided by an external generator; a programmer determines to which summing relay or amplifier the signal is provided.

In addition to all of the complexities inherent in accomplishing the above, the addition of summing relays or amplifiers provides one additional point where an error can develop. For these reasons, the most accurate and reliable system in which multiple pumps are involved uses a separate flowmeter and feeder for each measurement. This type of measurement and control system also results in each flow-measuring device operating over the flow range (usually only 3:1). If pumped flow signals are added or if the entire flow is passed through a single flowmeter, the required rangeability is extreme and accuracy is very poor. (For example, if three pumps of equal capacity are used, the required rangeability of the flow-measuring systems is approximately 9:1).

Controlling the chlorine feeders can be accomplished by continuous modulation or impulse duration. Neither system has any particular advantage over the other except where specific types of feeders are included in the evaluation.

Typical Control Instrumentation

Instrumentation for an offline storage and disinfection facility is selected on the premise that such a facility is often unmanned. Properly selected instruments will yield records that permit an operator to evaluate whether or not the equipment is performing correctly. Such a concept helps to eliminate the redundancy required by a completely unmanned facility. The occasional presence of personnel is required because of the size and complexity of the mechanical equipment, not because of the instrumentation. Facilities of the size normally required do not lend themselves to a completely unattended operation. The choice of operating media for instrumentation is influenced by the following conditions:

- The presence of high humidity and possibly corrosive vapors
- Discontinuous operation
- A lack of adequate maintenance.

Certain instrumentation, notably analytical measurements and particularly residual chlorine, requires special care to ensure that the equipment is protected from moisture and corrosion.

As the facility goes on line, the following sequences (as shown in Figure 59) occur:

- The increasing level is detected by LS-1; this in turn activates LA-1 and initiates the treatment function.
- The level in the wet well is measured and transmitted by LT-2 to level recorder controller LRC-2. The signal from LRC-2 controls the pump speed so as to hold the wet well levels within acceptable limits.
- The pumped flow from the wet well to the detention/sedimentation chamber is measured and transmitted by FT-3, recorded, and totalized by FR-3 and FQ-3, where a contact paces the chemical feeders.
- The chlorine residual is measured and transmitted by A-4 and AT-4 to recorder controller ARC-4. The controlling signal is fed back to the chemical feeder in order to trim its feed rate so as to achieve the desired chlorine residual.
- Additional alarm detectors can be added as required to any of the measuring or control signals.

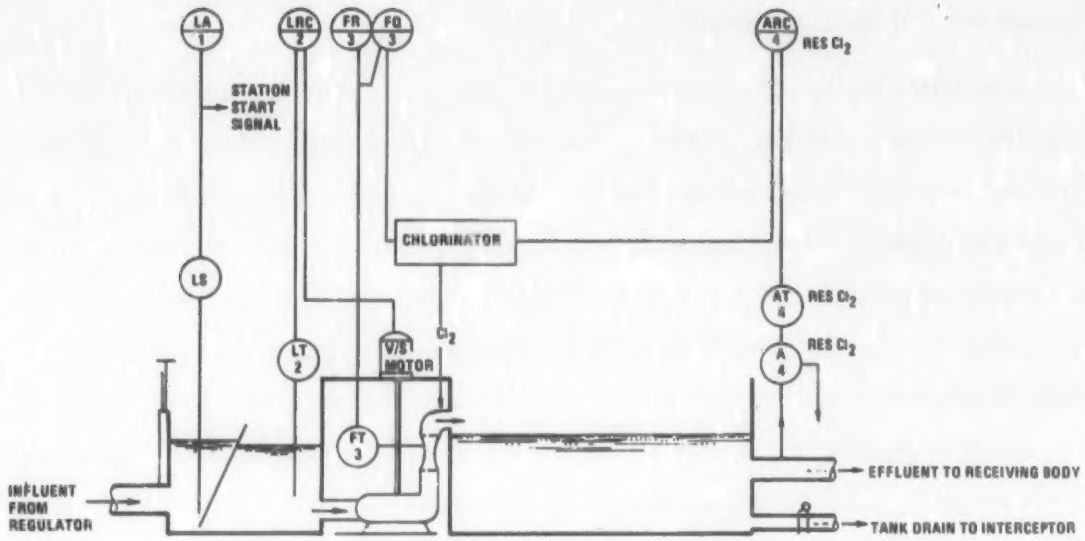

LA
1
LRC
2
FR
3
FQ
3
ARC
4
RES Cl2
STATION
START
SIGNAL
CHLORINATOR
LS
AT
4
RES Cl2
A
4
RES Cl2
LT
2
V/S
MOTOR
Cl2
FT
3
INFLUENT
FROM
REGULATOR
EFFLUENT TO RECEIVING BODY
TANK DRAIN TO INTERCEPTOR

Figure 59. Offline storage and disinfection plant instrumentation.

Since the pumping station incorporates multiple pumps, a sequencing and proportional level control system is required. As the level rises, successive pumps start and gradually increase in speed with the rising level. As a storm event comes to an end, the pumps decrease in speed and then drop out as the level decreases. The details of these control systems are a function of the type of pump drive and type of instrumentation.

The chlorination system is the other principal variable that can be incorporated into a control strategy. A number of problems, which are unique to this type of facility, interfere with complete design flexibility. Some of these are:

- Flow-measuring devices are usually quite difficult to adapt to these facilities.
- Chlorine feed equipment is expected to operate intermittently.
- Because these facilities are usually partially underground, the use of cylinder-stored liquid chlorine imposes a safety hazard to operating personnel.

- Chlorine (in the form of hypochlorite) may be in storage for extended periods of time with a resultant loss of strength.
- The monitoring and sampling equipment constitutes a serious maintenance problem because of intermittent operations.

Flow measurement is quite troublesome because design constraints usually result in long control weirs on either the inlet or outlet. During low flows the weirs are not properly ventilated, and the level change during high flows is small due to their length. Pipelines entering the treatment facility do not always flow fully, making application of either open-conduit or closed-conduit primary devices almost impossible. Wherever the plant inflow is delivered to the detention tanks by pump, conventional venturi-type devices or magnetic flowmeters can be used. Wherever large open channels are encountered, combining Pitot magnetic flowmeters and level devices provides a means of approximating the flow.

The essence of these problems is that storm-water chlorination on a flow-proportional basis is not usual or routine in comparison to experiences in sewage treatment plants. Although flow-proportional control is used commonly in sewage treatment plants, it may not be a viable choice for storm-water installations.

Residual control of chlorination is now desirable, since it directly controls the final variable and compensates for changes in chlorine demand. However, in highly transient flows serious time delays can occur between a residual chlorine measurement and an automatic correction. Another problem is maintaining the sampling system and residual measuring equipment in a ready state during the long periods between storm events, but this can be done by continuously circulating potable water through these units during shutdown.

The application of compound chlorine control assumes a viable means of measuring the station influent rate and residual. Unfortunately, the generally unsatisfactory flow arrangements for station influent seriously limit this technique. Although individual evaluation is essential in determining the practicability of compound loop control, it is used successfully in many chlorination and detention storm-water treatment facilities.

Table 12, which lists the advantages, benefits, confidence levels, and limitations of the chlorination control strategies, should guide the reader in selecting the most feasible chlorination control strategy.

Table 12. WET-WEATHER DETENTION AND CHLORINATION

Control Method	Benefits and Potential Savings	Advantages	Disadvantages
Chlorination in proportion to flow	Effective chlorination with minimum Cl_2 consumption	Simple and effective	Difficult to measure flow; does not compensate for any changes in chlorine demand
Residual chlorine feedback control	Effective chlorination with minimum Cl_2 consumption	Effective	Questionable value because of highly transient flows and irregular time lags in chlorine analyzer
Compound loop control (Residual chlorine feedback with fine chlorine trim)	Effective chlorination with minimum Cl_2 consumption	Very responsive to changes in both flow and demand	Requires a properly operating analyzer and chlorinator; very subject to malfunctioning

Instrumentation Maintenance

The philosophy governing operation and maintenance must consider the fact that this facility consists of batching or discontinuous processing operations. Most waste treatment facilities never completely shut down, once they start. But in this facility, the most critical time for the instrumentation is during startup—the instrumentation must work.

Accordingly the instrumentation must receive regular routine maintenance, even though the facility has not been activated for some time. Such routine maintenance should adhere to the following (approximate) schedule:

- Weekly—Manually synthesize a storm event and activate the analytical instrumentation (however, this is not necessary if the installation is actually operating on a once-a-week frequency)

- Monthly—Inspect and service each device; check the operation of the sampling systems
- Semiannually—Inspect and service the compressed air systems.

Normally a facility of this size is located in a metropolitan area of such magnitude that a central municipal wastewater-treatment facility already exists. In this case, arrangements should be made with the pertinent city department to have a competent instrument serviceman available, and he should be assigned to service the facility instruments on a regular basis.

COMBINED SEWER OVERFLOW SYSTEMS (HYPOTHETICAL LARGE DRAINAGE AREA OF 100,000 ACRES)

The combined sewer system servicing a large area usually accommodates storm flows by diverting the excess flow to adjacent watercourses. In the traditional system, the overflow occurs at regulator stations that are so arranged that, during dry weather and for flows up to two or three times dry-weather flows, they are diverted to an interceptor for subsequent delivery to the treatment plant. This arrangement requires a regulating device that operates so as to divert to adjacent watercourses all flows in excess of the interceptor capacity.

Regulators are generally designed as self-actuated devices. Historically they have not functioned very efficiently because of: 1) a lack of operating power inherent in self-actuated regulators, and 2) inadequate maintenance.

Even if these devices did operate with complete satisfaction, they still would not dramatically reduce the stream pollution caused by combined sewer overflows. It has been suggested that use of the storage capacity of the combined sewer system be made, thereby reducing the frequency and duration of overflows. Recent efforts in this area have examined and demonstrated the practicability of doing this (38, 39). However, such a proposal does involve a number of conditions that must be satisfied. These are as follows:

- The regulator devices must operate reliably.
- Some means for completely stopping the overflow must be provided.
- Monitoring a number of variables is required.
- A supervisory program for the operation of regulator stations must be available or developed from experience, math models, etc.

Satisfying the need for reliable regulator devices, as well as for some means of completely stopping the overflow, requires the use of externally powered gates, dams, etc. This requirement in turn can be met by using electric, hydraulic, or pneumatic actuators. A number of conventional sluice gates can also be adapted to this service.

Variables to Be Monitored

The use of the system storage capacity implies knowledge of hydrologic events occurring in the total drainage basin. Storms occurring in such remote areas may impose a heavy hydraulic load but be displaced in time. Knowledge of these events does permit anticipatory actions such as the dewatering of interceptors and the raising of overflow barriers. Local measurements, while necessary, do not permit such system preparation. These conditions require the monitoring of rainfall and sewer levels at areas remote from the point of regulation.

The regulation facility requires monitoring the following variables:

- The interceptor level
- The combined sewer level
- The equipment status (e.g., opened, closed, etc.).

It is commonly recognized that the flow should be monitored, but the obstacles to in-line storm water and combined flow measurement are almost insurmountable on both a physical and economic basis for the following reasons:

- Combined sewage contains large quantities of suspended and floating solids that make the maintenance of conventional flow-measuring devices a nearly continuous process.

- The application of standard designs of flow-measuring devices and incidental hydraulic requirements make each installation customized and seriously infringe on the adaptability of standard devices to existing installations.
- Devices such as magnetic flowmeters and sonic meters are costly.

As a result, the storage capacity and flows must be inferentially determined from level, and this requires an evaluation of applicable methods. The methods of level measurement that must be considered include the following:

- Float and cable (and variations)
- Capacitance
- Bubble tube.

The classic float and cable instrument is a poor choice for this service since it accumulates solids on the float, requires a direct mechanical connection to the readout or transmission device, and also requires extra maintenance when used for this type of service. The same is generally true of related devices that use a ball and arm.

A type of capacitance device that may be applicable for flow measurements in sewers has one plate located above the flow and parallel to the surface of the flow. The flow itself is the second plate through a ground connection since the plate areas are constant and capacitance is a measure of the distance between the plates. There are just two practical problems: the relationship between the level and the capacitance is nonlinear, and the span is small.

The classic bubble tube appears to be the most practical level detection device for this application. It is completely flexible, it is a static-measuring system, and it is readily adaptable to all transducer types. With a little imagination, this tube can be adapted to practically any installation (either proposed or existing). It is quite possible to operate this device from bottled compressed air, nitrogen, CO_2, or from a small compressor.

Transducers and Transmission

The transducing of measurements and equipment status may be accomplished in several ways. In the case of equipment status (e.g., a valve opening or closing, a motor running, etc.), the presence or absence of a signal or two different states of a signal suffice. Wherever a continuously monitored and modulated signal is required, there are two common methods of signal conditioning:

- Modulating the signal to the impulse duration
- Modulating the signal to the variable-rate frequency shifts.

These signals in turn are usually converted to audio tone signals for transmission by leased telephone circuit or microwave. The choice, which is a function of a number of considerations, is largely related to communication link requirements.

Regulator Station Automation

The regulator station requires the following measurements for monitoring and controlling combined sewage overflows:

- The level in the combined sewer upstream of the control device (e.g., the sluice gate, etc.)
- The level in the interceptor
- The level in the combined sewer downstream of the control device (e.g., the sluice gate, etc.)
- The status of control elements [e.g., the sluice gate (whether opened or closed), etc.]
- The station power supply and security status.

In turn, these measurements require the following equipment items:

- Three bubble tubes, with associated level transmitters and a compressed air supply
- One combined sewer control element (e.g., the sluice gate, fabridam, etc.)

- One interceptor control element (e.g., the sluice gate, valve, etc.)
- Signal conditioning devices (as required)
- Communication equipment
- Local control stations.

While many combinations of arrangement and equipment choice are possible, the actual arrangement and equipment choice are governed largely by existing conditions and so can be depicted only in a general manner. A schematic diagram representing a typical arrangement is shown in Figure 60.

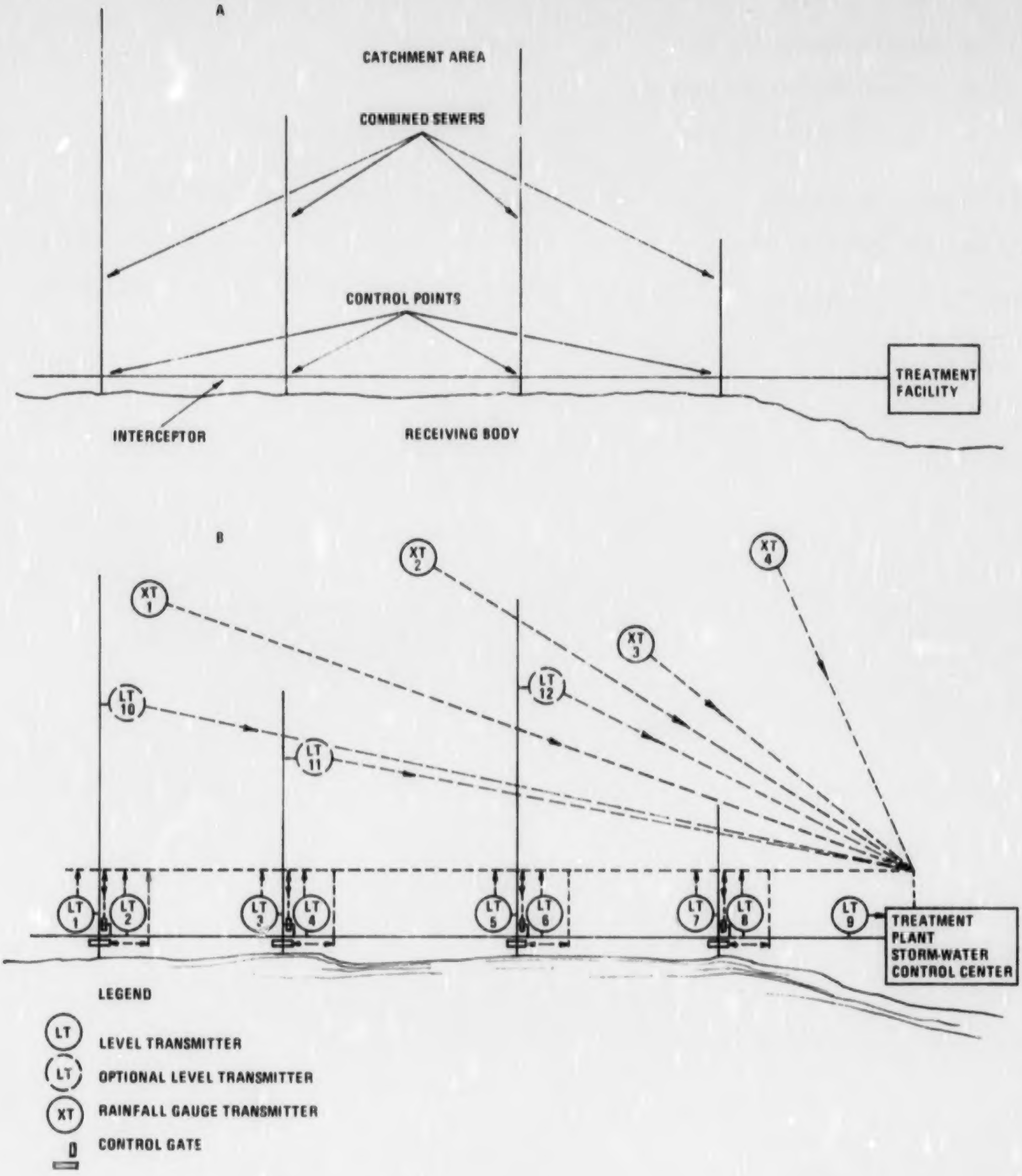

A
CATCHMENT AREA
COMBINED SEWERS
CONTROL POINTS
TREATMENT FACILITY
INTERCEPTOR
RECEIVING BODY
B
XT 1
XT 2
XT 3
XT 4
LT 10
LT 11
LT 12
LT 1
LT 2
LT 3
LT 4
LT 5
LT 6
LT 7
LT 8
LT 9
TREATMENT PLANT STORM-WATER CONTROL CENTER
LEGEND
LT
LEVEL TRANSMITTER
LT
OPTIONAL LEVEL TRANSMITTER
XT
RAINFALL GAUGE TRANSMITTER
CONTROL GATE

Figure 60. Combined sewer overflow control.

SECTION VIII

COST/BENEFIT ANALYSIS

INTRODUCTION

Two objectives of this report are to show the economic feasibility of automation as compared to manual operation and to compare the economics of competing control strategies where appropiate. Cost/benefit analysis such as payback period appears suited to accomplish these objectives. This technique is described by Peters and Timmerhaus (48) for industrial applications. Payback period has the advantage that it allows a business to determine the length of time required to recoup its investment based on the savings and profits from the installation and thereby, with one simple figure, management can make a reasonable decision as to the effectiveness of the investment. Since these cost estimates are being used for comparative purposes, it was not necessary to obtain the best accuracy and therefore the data is limited to preliminary design estimates.

Since municipal wastewater treatment facilities do not, in general, operate to make a profit, the payback period as applied to municipalities has been modified to reflect the time necessary to pay for the installation of all equipment for automation based on the savings from that automation. To properly explain this concept, the following example is used. When a cost/benefit analysis comparing an automated versus a manual method was performed on a municipal wastewater treatment plant, a potential gross savings of $50,000 resulted because of automation. The automation requires the use of an analyzer, controller and final control element. The cost and life expectancy of these instruments, for our purposes, will be assumed as a:

- $10,000 instrument value with 5 years life expectancy
- $50,000 instrument value with 10 years life expectancy
- $10,000 instrument value with 15 years life expectancy.

In addition to the capital investment needed, there would be yearly operating and maintenance expenses. This yearly cost in our example is estimated to be 800 manhours or, at $10 a manhour, $8000/year. If one assumes an interest rate of 6% for municipal loans with 10 years or longer duration and 8% for 5 years or shorter duration, the list of assumptions is now sufficient to illustrate the required steps in a cost/benefit analysis via the total annual cost technique.

First, one should convert all the dollar figures into annualized values. The benefits are already in that form ($50,000/year) so that only the capital cost figures need to be converted. This is done as follows (49):

total annual cost = annual capital recovery cost + annual labor cost + annual replacement part cost*

$$= \frac{P\,(1+i)^N i}{(1+i)^N - 1} + LC + RC$$

where:

P = amount of money needed at present time

i = prevailing interest rate, fraction

N = useful life of equipment, years

LC = annual labor cost, $/year

RC = annual cost of replacement parts, $/year

Substituting values in the above equation results in an annual cost of $18,340.

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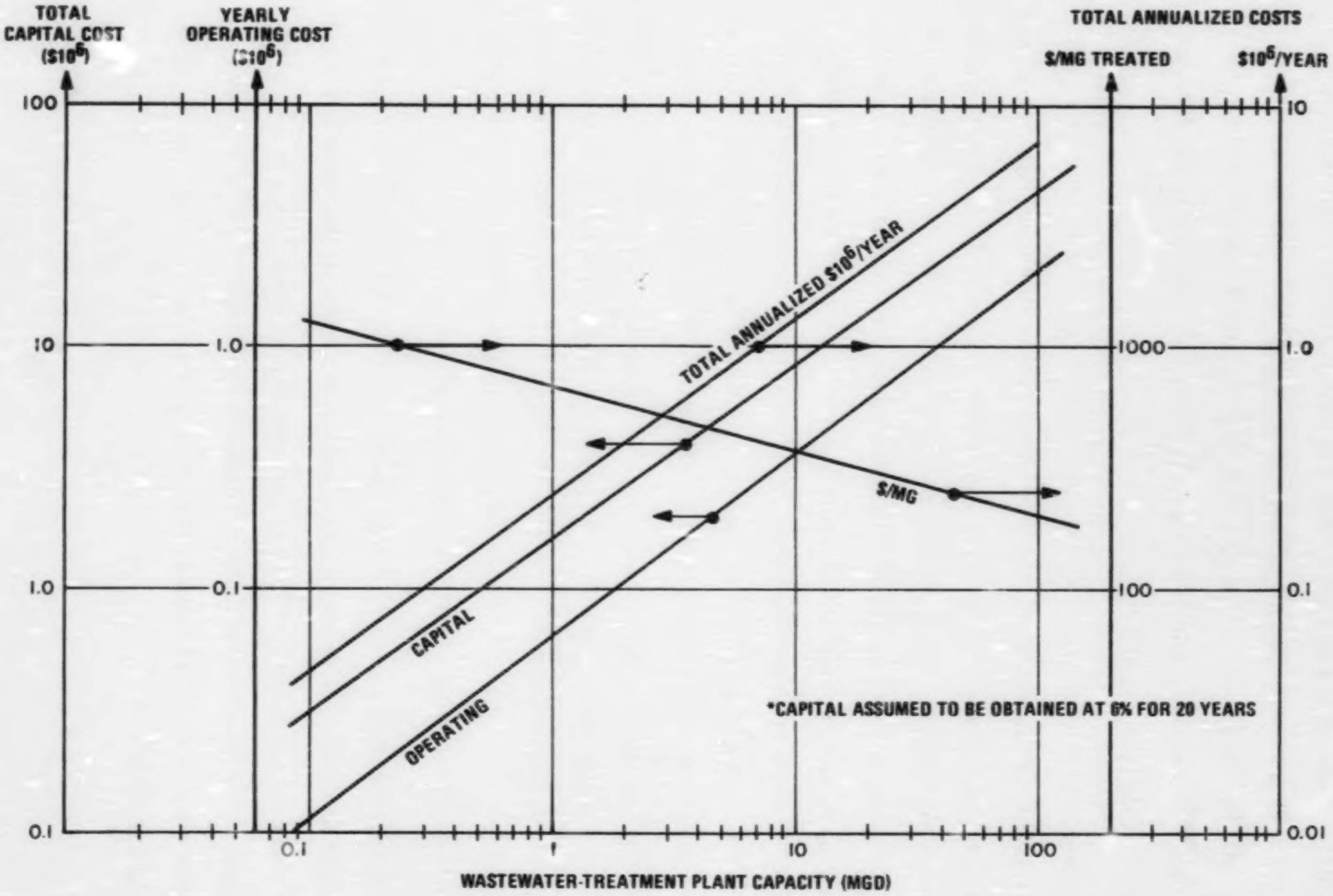
*The equipment is assumed to be incapable of being salvaged after its useful life.

The net yearly savings is the gross yearly savings less the annual cost or $50,000 - 18,340 = $31,660. This represents the yearly savings of 45% in terms of added instrumentation cost ($31,660/70,000 = 0.45). Another way of expressing this relationship is by saying that the payback period for this amount of instrumentation is 2.2 years. This data may be summarized as: total capital cost/yearly net savings = $70,000/(50,000 - 18,340) = 2.2 years.

Since large amounts of capital are needed for the construction and upgrading of municipal and industrial wastewater plants, accurate and realistic economic analyses are essential for an intelligent allocation of resources. Specifically, a return on an investment and a payback period are the traditional financial criteria used as decision-making aids in selecting the most desirable investments. Although the particular criteria depend upon the economic climate (for instance, the interest rate or availability of funds), it shall be assumed in this report that current payback periods of approximately two and one-half years or less are needed to justify the installation of optional instruments and automatic control equipment. In some cases where the equipment becomes integral to operation of the plant (e.g., computers) or where the equipment is expected to have a long life, management may then find payback periods in excess of two and one-half years justifiable. When some form of instrumentation or automatic control is essential and several competitive approaches are available, payback periods can be used to select the most desirable alternative.

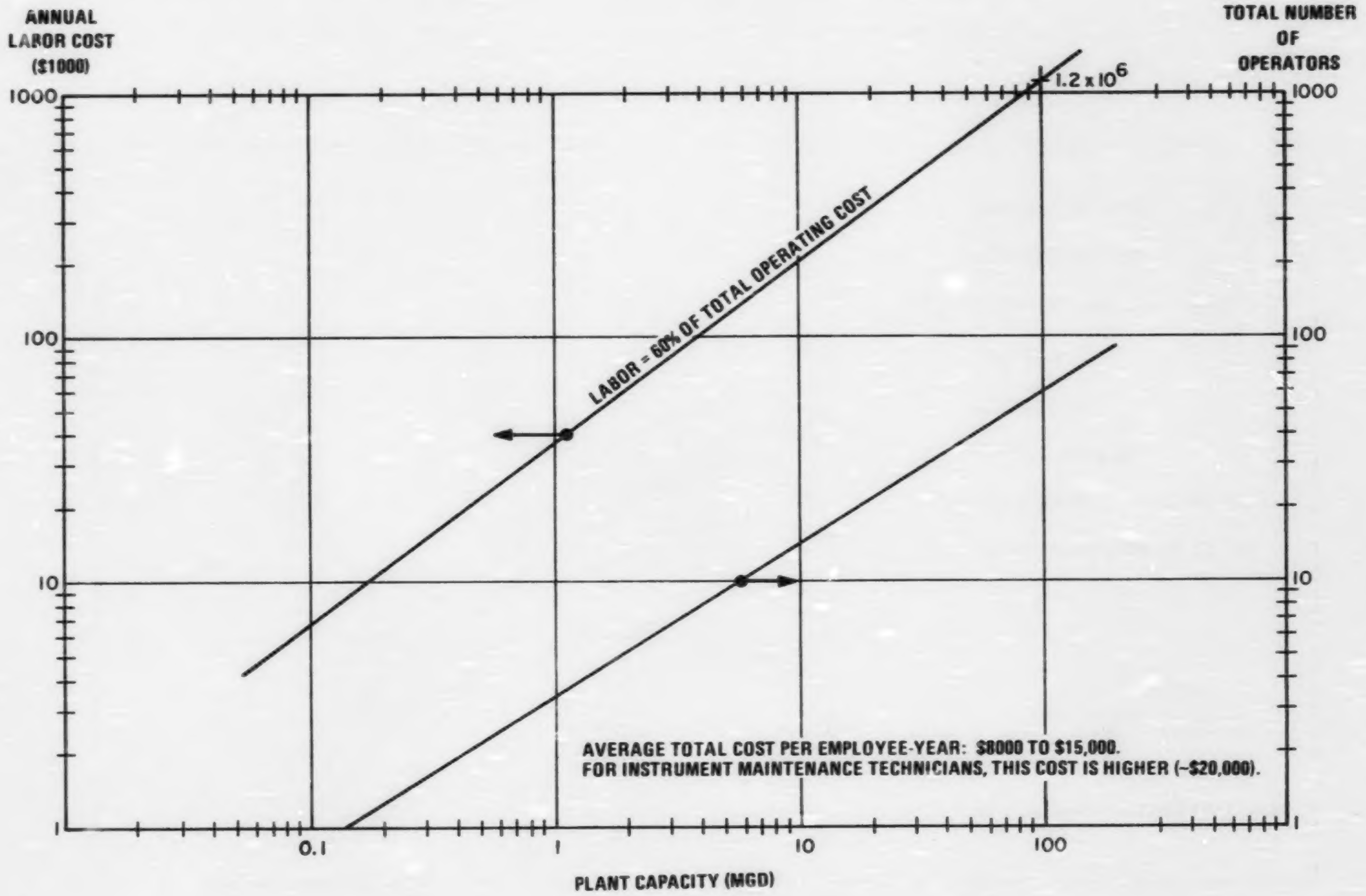
This section details the general methodology of the cost/benefit analysis and shows the application to the alternate control strategy selection for some of the more important wastewater-treatment processes. These techniques can be readily adapted to other treatment processes and instrumentation schemes by the reader. Also, the cost/benefit analyses may be easily updated to reflect new costs and interest rates. (See Figures 61, 62 and 63.)



*CAPITAL ASSUMED TO BE OBTAINED AT 6% FOR 20 YEARS

NOTE: The type of plant described is a multistage biological process providing the following reductions: OD–95%, SS–100%, P–100%, and N–85%.

Figure 61. Total annualized wastewater-plant costs.



AVERAGE TOTAL COST PER EMPLOYEE-YEAR: $8000 TO $15,000.
FOR INSTRUMENT MAINTENANCE TECHNICIANS, THIS COST IS HIGHER (~$20,000).

Figure 62. Wastewater-plant labor costs.

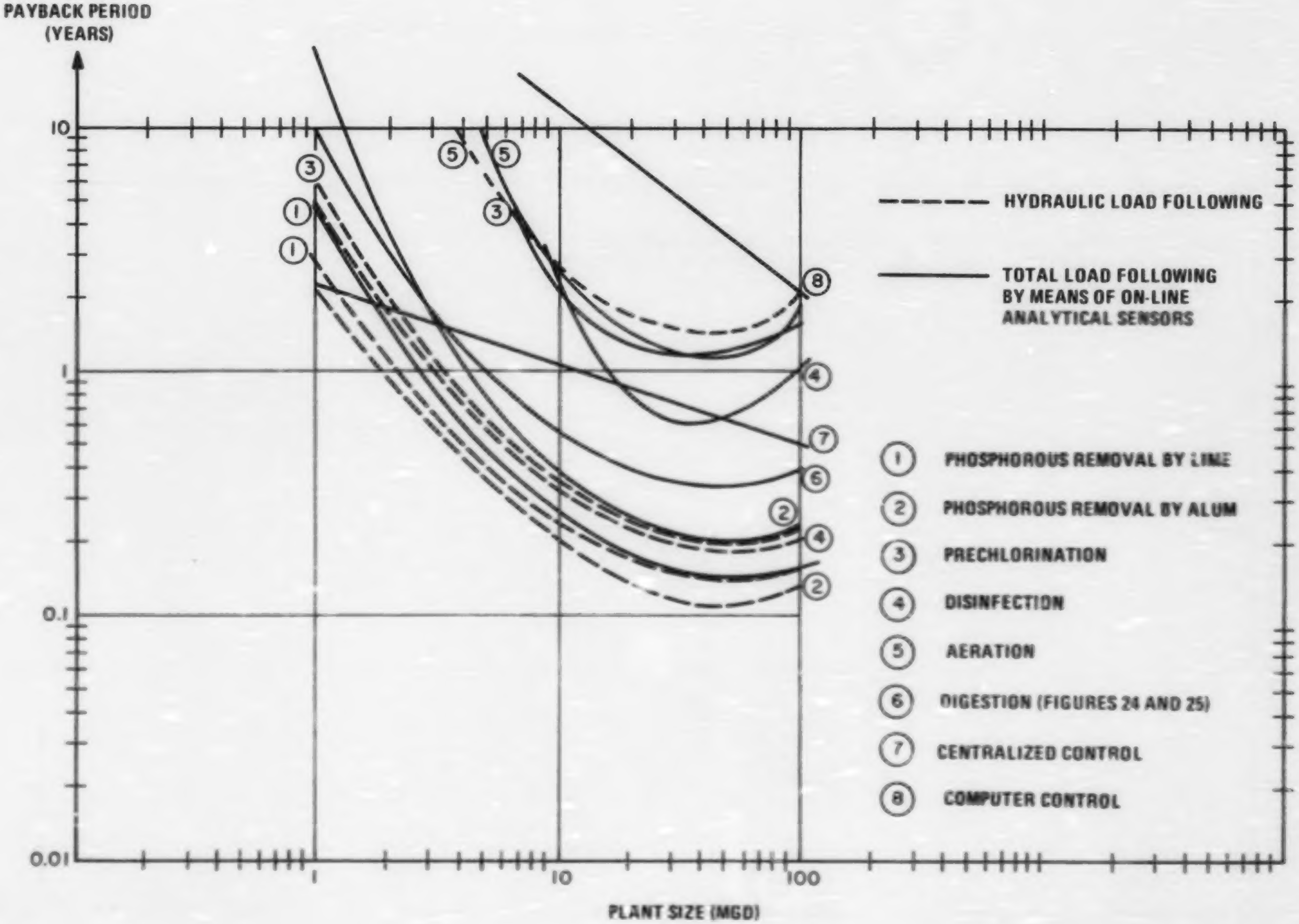


Figure 63. Automatic control payback periods vs plant size for typical municipal wastewater treatment.

Basis for Economic Comparisons

Several alternate control systems are evaluated for each of the main unit operations in industrial or municipal wastewater-treatment plants. Some of these control schemes are superior to the manual (uncontrolled) mode of operation in that they can guarantee consistent and acceptable effluent quality, whereas others can demonstrate an economic advantage because of the resulting increase in equipment life and the reduction in:

- Equipment size
- Operating labor
- Maintenance requirements
- Consumption of chemicals
- Consumption of electric power
- Quantity of sludge produced.

The balance between costs and benefits is also affected by the size of the plant, with the larger operations being in a better position to justify the initial cost of a higher level of automation. There is no dollar value attached to the consistent meeting of effluent quality regulations by automation devices, nor is there a cost penalty assigned to the inconsistent performance of manually operated plants. These are treated as extra bonuses or penalties that are difficult to be quantitatively evaluated. It is difficult to quantify any economic advantages gained because of increases in effluent quality.

Throughout this section, the dollar values used are 1974 first-quarter dollars, which correspond to 362 on the Marshall & Stevens scale and to 151 on the Chemical Engineering cost index. Unless otherwise noted, the plant lives are assumed to be 25 years. The interest rates are assumed to be 6% for instruments with a life expectancy above 10 years, and 8% for instruments with a life expectancy below 10 years. Chemical coats used in the cost benefit analyses are representative averages. No provisions have been made with regard to volume, method of shipment, storage, or plant location cost differentials. As a result, the methodology in this section should be applied as a guideline.

Evaluation of Cost Benefits

Table 13 provides the approximate operating and capital costs of wastewater-treatment plants (50) together with their total numbers in the U.S. Usually less than 4% of the listed capital cost is spent on instrumentation, and more than half of the operating cost is spent on labor. A recent nationwide survey of wastewater-treatment facilities found that most secondary plants spend about 3.3% of their capital cost on instrumentation (1). Finally, treatment plant capital and operating cost data, expressed as a function of plant size (as in Figure 61), clearly shows the economy of scale. This leads to a popular misconception which is that only large plants (i.e., 50 mgd) can afford the luxury of instrumentation and automatic control budgets. However, the higher unit treatment costs of smaller sized plants may make it feasible for medium-sized plants (i.e., 5 mgd) to reap even greater benefits from instrumentation.

Table 13. APPROXIMATE COSTS OF WASTEWATER-TREATMENT PLANTS

Costs	Plant Capacity (mgd)		
	10	50	100
Yearly operating costs ($\$10^6$/year)*	0.3	1.0	1.5
Capital cost ($\$10^6$)	8.0	25.0	45.0
Number of domestic municipal and industrial plants in this size range	500.0	150.0	100.0

*According to Reference 51, the components of total operating and maintenance costs in the 1965 to 1968 period in municipal plants were: 1) labor—60%, 2) electricity—14%, 3) chemicals—4%, and 4) others—22%.

The installed cost of wastewater-treatment equipment is substantial and well documented (52). The use of state-of-the-art instrumentation may reduce the quantity and/or size of these devices by guaranteeing their optimized utilization and by eliminating their intermittent use. One example of quantity reduction would be the continuous—instead of intermittent—use of vacuum filters, while one example of size reduction would be the feedforward control of pH, which allows for a substantial reduction in the size of the neutralization tank. Similarly, the size of pumps can be reduced and their life increased if they are operated continuously instead of on and off.

Well-designed control systems make the various pieces of equipment work only as hard as necessary. This (in the case of aerators, for example) can result in a 10% increase in equipment life.

Automation can also eliminate the misoperation-type of failures due to human error. Burning-out motors, flooding, or running the equipment dry are all prevented by well-designed instrumentation, which pays off in lower maintenance expenses, less equipment downtime, and higher overall efficiency.

Compared to all other operating expenses, labor costs are the highest in wastewater-treatment plants. If the decision-making process (leading up to a manual adjustment made by an operator) can be reproduced by instrumentation, the automated system will perform more consistently, accurately, and reliably, in addition to relieving the operator of that task. While increased levels of instrumentation will reduce the operating labor costs of the plant, more sophisticated instruments and analyzers would require a higher level of maintenance attention. The probable overall saving is treated as an additional benefit from automation and is not included in the quantitative cost/benefit analysis for the various unit operations. However, the potential reductions in operating labor costs via the labor-saving role of centralized and computerized control are significant and are discussed subsequently in this section in detail.

Table 14 lists some of the unit costs and dosages of chemical additives (53). On an average, chemicals account for less than 10% of the total operating costs. Automation can substantially reduce the total use of chemicals by continuously monitoring and matching only the actual demand. However, only a small number of chemical addition loops are actually controlled on a demand basis other than flow proportional. In fact, the field survey of user experiences (1) observed no cases of significant chemical

Table 14. COST OF CHEMICAL ADDITIVES*

Cost/Ton ($)		Hydrated Lime	Ferric Chloride	Chlorine
		20	100	100 **
Typical Dose (mg/liter)		250	100	15
Corresponding Yearly Cost ($)	1 MGD Plant	8,000	16,000	2,500
	10 MGD Plant	80,000	160,000	25,000
	100 MGD Plant	800,000	1,600,000	250,000

*Other chemical costs/ton ($): alum $80 (costs are based on a liquid alum addition at $80/ton of 22% dry Al_2O_3), ammonia $70, caustic (100%) $140, sulfuric acid (100%) $40, HCl (36%) $40, SO_2 liquid $75, and polymers $1000 to $2000.

**Cost does not include shipping or cylinder rental charges.

savings by means of demand control with the exception of chlorine. The increased popularity of physical/chemical treatment and chemical advanced wastewater treatment will undoubtedly increase the significance of chemical saving control systems. In plants where chlorination, flocculation, coagulation, phosphorous removal, and/or neutralization are also practiced, the resulting savings can be very substantial.

BASIC ASSUMPTIONS COMMON TO ALL UNIT OPERATIONS

Load Following

All wastewater-treatment facilities experience cyclic (diurnal) variations in their hydraulic and organic loads. In order to lessen the use of chemical additives, the best control method is to ratio the additive flow to the pollutant load. A halfway measure ratios the rate at which the additive is charged to the volumetric flow rate of the incoming wastewater, while the least desirable technique charges the additive at a constant rate.

In order to evaluate the benefits of hydraulic load following, it is necessary to approximate the amount of chemical additives that will be conserved through use of this method. The percentage of these chemicals is estimated by a comparison with the constant feed method, where the addition rate is set for the maximum load and is left unaltered during the day. The relationship between the peak and average flows is as follows (54):

$$F_p = 1.84\,(F_a)^{0.92}$$

where:

F_p = peak flow

F_a = average flow

The peak-to-average ratio decreases with increasing flows. These are listed in Table 15 for 1, 5, 10, 50, and 100 mgd average flows.

Table 15. SAVINGS IN CHEMICAL ADDITIVE USE BY PRACTICE OF LOAD FOLLOWING

Average Plant Flow (F_a) (mgd)	Hydraulic Load Following		Total Load Following	
	Peak-to-Average Ratio of Flow ($K = F_p/F_a$)	Percent of Savings Over Constant Rate Charging ($\% = \frac{K-1}{K}$)	Peak-to-Average Ratio of Chemical Additive Use ($A = 0.75K^2 + 0.25K = F_pC_p/F_aC_a$)	Percent of Savings Over Constant Rate Charging ($\frac{A-1}{A}$)
1	1.84	46	3.00	67
5	1.62	38	2.38	58
10	1.53	35	2.13	53
50	1.34	24	1.68	40
100	1.27	21	1.53	35
500	1.18	15	1.34	26

If the daily hydraulic load variation is assumed to be sinusoidal (6) around the average flows, then setting the chemical addition rate continuously at a level corresponding to the plant's peak flow will result in unnecessary overcharging of the chemical additive by the ratio of F_p/F_a, as shown in Figure 64 and Table 15.

The concentration of a wastewater stream also varies according to a diurnal schedule (6). This variation can also be considered sinusoidal in character and is in phase with the hydraulic load variation. In other words, at low hydraulic loads the concentration is also likely to be low, and at peak volumetric flows the pollutant concentration is also likely to be high.

The savings in the consumption of chemical additives, therefore, will be even greater when the feature of pollutant concentration load following is added on top of the hydraulic flow ratioing. This amplification effect can be made quantitative by assuming that the pollutant concentration variation equals 75% of the hydraulic load variation (55). This can be expressed as:

$$C_p = (0.75(K-1) + 1)\ Ca$$

where:

C_p = peak concentration

Ca = average concentration

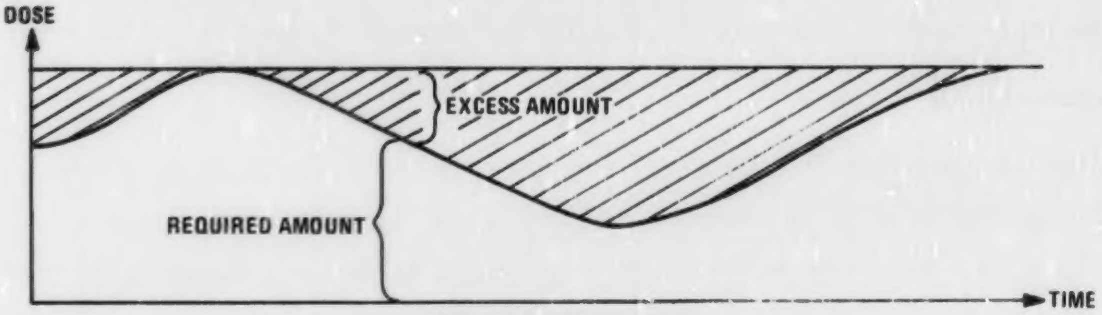
$K = F_p/F_a$

If the chemical additive was charged at a constant rate, its flow would correspond to the peak values of both the pollutant concentration and hydraulic load:

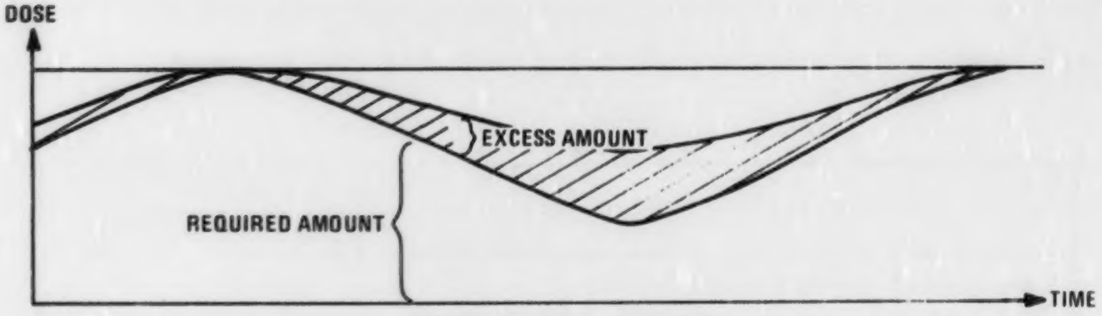
$$\text{constant rate setting} \simeq (F_p)\ (C_p) = (KF_a)\ (0.75(K-1) + 1)\ Ca$$

If it is assumed that, with the total load following, the rate of chemical additive usage corresponds to $F_a Ca$, then the peak-to-average chemical use ratio is:

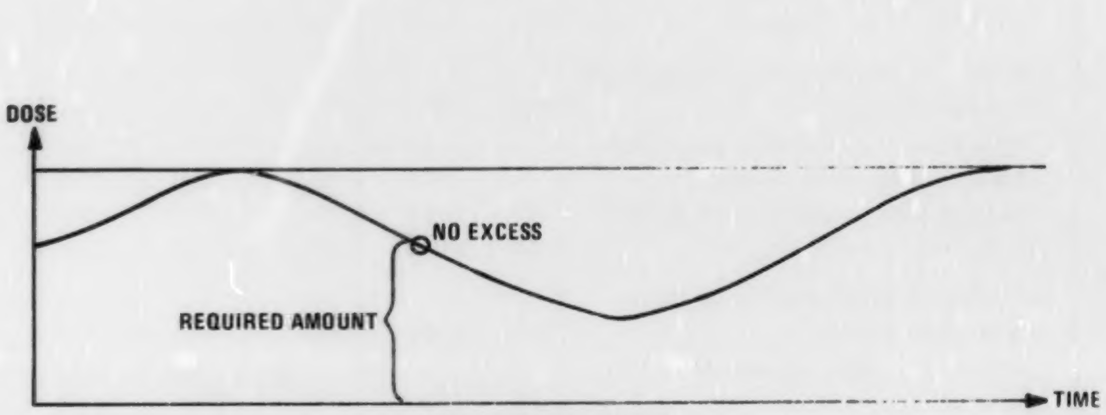
$$F_p Cp/F_a Ca = 0.75K^2 + 0.25K$$



CONSTANT RATE ADDITION



HYDRAULIC LOAD FOLLOWING



TOTAL LOAD FOLLOWING

Figure 64. Various methods of chemical additive changing.

The percentage savings accomplished by the total load following is provided in Table 15, while the three methods of chemical additive charging all illustrated in Figure 64.

Chemical Additive Doses

After approximating the percentage savings potential of the various control strategies, the next logical step is to determine the quantities of chemicals to which these percentages apply. Table 16 lists some of the more important unit operations, together with typical dosages of chemical additives (assuming a constant rate of charging) and their daily cost for various plant sizes.

The daily cost of makeup is based on the cost data in Table 14, and the daily cost of chemical sludge disposal is estimated on the basis of \$5/wet ton, with alum treatment

Table 16. CHEMICAL ADDITIVE DOSES

Unit Operations	Plant Size (mgd)				
	1	5	10	50	100
Odor control (prechlorination) (57)					
Cl_2 dose (ppm)	5	5	5	5	5
Daily Cl_2 usage (lb/day)	42	210	420	2100	4200
Daily Cl_2 cost (\$/day)	5	25	50	250	500
Phosphorus precipitation by lime (56)					
Lime dose (ppm)	500	500	500	500	500
Daily lime consumption (tons/day)	2.1	10.5	21	105	210
Daily makeup cost (\$/day)	42	210	420	2100	4200
Daily sludge disposal cost (\$/day)	15	75	150	750	1500
Daily total cost (\$/day)	57	285	570	2850	5700
Phosphorus precipitation by alum (58,59)					
Alum dose (ppm)	200	200	200	200	200
Daily alum consumption (tons/day)	0.83	4.15	8.3	41.5	83
Daily makeup cost (\$/day)	66	330	660	3300	6600
Daily sludge disposal cost (\$/day)	1.25	6.25	12.50	62	125
Daily total cost (\$/day)	67.25	336.25	672.50	3362	6725
Disinfection via chlorination (57)					
Cl_2 dose (ppm)	15	15	15	15	15
Daily Cl_2 usage (lb/day)	126	630	1260	6300	12,600
Daily Cl_2 cost (\$/day)	15	75	150	750	1500

resulting in 0.25 ton of wet chemical sludge after each mg treated, and lime producing 3 tons of wet chemical sludge per each mg treated (56).

One can use the data in Table 16 to calculate quantitative costs/benefit. For example, to calculate the potential daily savings in operating a 50 mgd phosphate precipitation unit that uses alum by converting its control strategy from hydraulic load following to total load following, Tables 15 and 16 contain the data required. In Table 15, the percentage of alum saved is found to be 40 - 24 = 16%. In Table 16, the daily cost savings is found to be (0.16) (3,362) = $540. In figuring a cost/benefit analysis, it is necessary to balance this savings against the cost of added instrumentation.

Parallel Trains of Equipment

Because the instrumentation cost is a function of the number of parallel trains of equipment rather than the wastewater-treatment plant capacity, some assumptions are desirable in this respect. Table 17 lists some of the major pieces of equipment and the probable number of parallel units that are likely to exist in plants of various sizes. These quantities form the basis of the cost/benefit analysis of various control strategies (60).

In Table 17 it can be noted that the quantity of instrumentation (not the size of pipeline items) is about the same for 1, 5, and 10 mgd plants. Therefore, wherever the benefits of automation are proportional to plant capacity, it will be much easier to justify a higher level of automation for a 10 mgd plant than for a 1 mgd one.

Table 17. NUMBER OF PARALLEL UNITS

Unit Operations	Plant Capacity (mgd)					
	1	5	10	50	100	500
Primary clarifier	2	2	2	4	8	20
Aeration tank	2	2	2	4	8	20
Secondary clarifier	2	2	2	4	8	20
Sludge dewatering (vacuum filter)	2	2	2	4	8	20
Sludge digester	2	2	2	4	8	20

Assumptions on Sensors and Final Control Elements

In order to establish a common basis for comparing the costs of more sophisticated control systems to manual (or operator-controlled) installations, it is essential to define the types of instruments that must be added for automatic operation. It is also necessary to make some assumptions about the types of instruments that already exist in a manually operated plant, since the cost of these devices does not enter into a relative economic evaluation of the various control strategies. Such instruments will be distinguished on the control system drawings by two concentric circles.

A transmitter signal will be assumed to exist and be freely available for all control strategies. This assumption applies to all flow-ratioed additive control strategies, including deodorizing, neutralizing, precipitating, coagulating, flocculating, or disinfecting control systems.

Variable speed metering pumps and adjustable chlorinators will be assumed to be common to all control strategies. When it is sufficient to use such indirect flow data as chlorinator or metering pump setpoints, this information will be assumed to be freely available to all control strategies.

When direct and independent flow measurements are required, this will be assumed to be an extra-cost item. For example, if it is desirable to install a magnetic flowmeter to detect the flow rate of lime slurry, this will be treated as an additional-cost item for the particular control strategy involved.

A transmitter signal that represents the individual airflows to each aeration tank will be assumed to be freely available.

When sludge flow information is obtained indirectly from pump speeds, this data will be assumed to be freely available. When direct flow data is required for the control strategy, the cost of the sensor (such as a magnetic flowmeter) will be assumed to be extra. All magnetic flowmeters in sludge service will be assumed to be 6 inches in size and mounted in 8-inch diameter pipes (unless otherwise noted). For sludge service the magnetic flowmeters will be estimated to be uncalibrated weather-proof units with fiberglass lining and ultrasonic cleaners. The 1974 uninstalled costs that follow include the cable and the general-purpose converter required to generate a 3 to 15 psig output signal: 1) four inches $3130, 2) six inches $3380, and 3) eight inches $3900. Unless otherwise noted, the 6-inch size will be assumed to be applicable at a rounded-off unit price of $3500. Finally, the digester gas flow rate will be considered an extra-cost item for the control strategies needing this information.

When the final control element is a pump or a feeder, it is assumed to exist and be freely available; when it is a valve, it is assumed to be an extra-cost item.

For sludge service the control valves will be estimated as ductile iron Veebal valves. Their 1974 uninstalled costs are as follows: 1) four inches: $865, 2) six inches: $1005, and 3) eight inches: $1400. Unless otherwise noted in this text, the 6-inch size will be assumed to be applicable at a rounded-off unit price of $1000. This is the right size for an 8-inch sludge pipe installation (61).

If a control strategy requires measuring the interface between the sludge and effluent or detecting the percentage of solids in a flowing stream, the sensors required will be treated as extra-cost items.

All chemical analyzers required by a particular control strategy will be considered extra-cost items.

Cost of Instrument Maintenance

The following assumptions were used to arrive at an hourly rate for instrument repair, tuning, and maintenance:

- Maintenance is performed by in-house technicians.
- The base salary (and corresponding skill) of these technicians does not exceed the $7/hour rate.

- The level of benefits and the ratio of nonproductive overhead are those of a typical municipal operation.
- There is a scheduled preventive-maintenance program in effect to guarantee that the idle time of maintenance personnel does not exceed 10%.

Based on these assumptions, a $10/hour rate has been used throughout this section. Table 18 lists the estimated yearly maintenance requirements of the various types of instruments (1, 62) based on actual service in municipal wastewater-treatment plants and industrial facilities.

In small plants, it may be more economical to rely on contract instrument maintenance at approximately $25/hour rather than to provide a full-time in-house maintenance capability.

For the purposes of this document, the instruments are grouped into three classes (refer to Table 18 for specific values) (1, 62):

- Class I—Over 15 years life expectancy
- Class II—Ten years (or a 5 to 15 range) life expectancy
- Class III—Under 5 years life expectancy.

REVIEW OF UNIT OPERATIONS

For phosphate precipitation through lime addition the cost of additional instruments needed for hydraulic and total load following will be compared to the economic benefits of reduced lime usage and the reduced chemical sludge disposal cost.

When lime is charged at a fixed rate corresponding to the maximum flow rate and the maximum pollutant concentration during the daily cycle, it results in the maximum expense in lime makeup, CO_2 use and sludge disposal. The yearly total cost for various plant sizes (Table 16) is as follows:

Table 18. EXPECTED LIFE AND YEARLY MAINTENANCE REQUIREMENTS BY TYPE OF INSTRUMENT

Instrument Type	Maintenance (hours/year)	Lifespan* (years)
Panel-mounted devices		
Annunciators	2	15
Controllers	8	10
Converters	10	10
Indicators	4	15
Programmers	20	5
Recorders	8	10
Switches	2	15
Final control elements		
Position (control valves)	12	10
On-off valves, pumps	4	15
Variable-speed feeders, pumps	16	10
Transmitters		
Flow	8	15
Level	6	15
Others	10	10
Pressure	4	15
Temperature	6	15
Analyzers		
CH_4	50	5
Chromatograph	150	5
COD, TOC, TC	150	5
Combustibles	25	5
Conductivity	60	10
CO_2	25	10
DO	60	5
NDIR	75	5
Nuclear sludge density	50	10
O_2 in gas	40	5
pH and ORP	50	5
Phosphate	150	5
Refractive index	40	5
Residual chlorine	140	5
Respirometer, BOD	150	5
Turbidity	60	10
UV	60	5
Sensors		
Magnetic	12	15
Orifice	5	15
Position (displaced gas)	50	5
Position (displaced liquid)	12	10
Propellors	10	10
Venturi	20	15
Weir, flume	2	15
Level		
Bubbler	8	15
Capacitance	6	10
d/p	5	15
Float and cable	60	15
Nuclear	10	10
Pressure		
Bourdon	2	15
d/p	4	15
Temperature		
Bimetallic	4	15
Filled	6	10
TC, RTD	8	15
Weighing systems	60	10
Position (sluice gates)	30	5
Speed	10	5

*5 means 0 to 5, 10 means 5 to 15, and 15 implies 15 or more.

1 mgd	$ 20,500
5 mgd	$ 102,500
10 mgd	$ 205,000
50 mgd	$1,025,000
100 mgd	$2,050,000

The potential total savings resulting from the application of the hydraulic load following and total load following control strategies is given in Table 19 (multipliers from Table 15).

Table 19. BENEFITS OF HYDRAULIC LOAD FOLLOWING IN PHOSPHORUS REMOVAL VIA LIME ADDITION

Plant Size (mgd)	Savings Through Hydraulic Load Following ($/year)	Savings Through Total Load Following ($/year)
1	(0.46) (20,500) = 9,400	(0.67) (20,500) = 13,700
5	(0.38) (102,500) = 39,000	(0.58) (102,500) = 59,400
10	(0.35) (205,000) = 72,000	(0.53) (205,000) = 108,500
50	(0.24) (1,025,000) = 245,000	(0.40) (1,025,000) = 410,000
100	(0.21) (2,050,000) = 430,000	(0.35) (2,050,000) = 718,000

Figure 65 describes the instruments required for hydraulic load following. The corresponding capital and operating costs (63) are contained in Table 20. The annualized cost of one unit of this control system (64) based on a 6% interest rate, is $1695 for a useful life of 10 and 15 years as shown in Table 20.

It is assumed that one of these systems will be required for each primary clarifier in the plant. Therefore, based on Table 17, the annualized cost as a function of plant capacity is as follows:

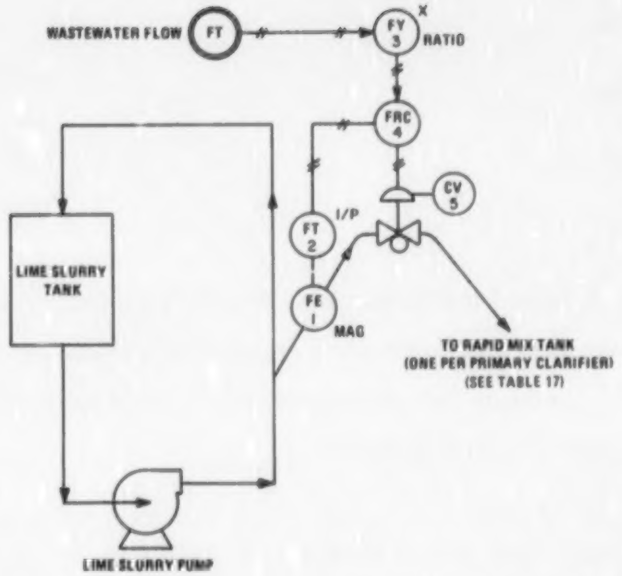

WASTEWATER FLOW
FT
FY 3
X
RATIO
FRC 4
CV 5
FT 2
I/P
FE 1
MAG
LIME SLURRY TANK
TO RAPID MIX TANK (ONE PER PRIMARY CLARIFIER) (SEE TABLE 17)
LIME SLURRY PUMP

Figure 65. Phosphorus precipitation by lime addition in hydraulic load following control mode.

Table 20. CAPITAL AND OPERATING COSTS OF HYDRAULIC LOAD FOLLOWING CONTROLS

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
10 years depreciation period		
FT-2	1200	10
FY-3	380	10
FRC-4	760	16
CV-5	720	12
	3060	
15 years depreciation period		
FE-1	1400	12
Installation materials	300	
Control panel section	350	15
Engineering and design	1400	
Installation and startup labor	1700	
	5150	75 hours/year
Total installed cost per loop	8210	
Total annual operating cost per loop		$750

1 mgd	$ 3,390
10 mgd	$ 3,390
50 mgd	$ 6,780
100 mgd	$13,560

Considering the yearly cost of instrumentation listed above and the resulting yearly savings (given in Table 19), this control system can be justified for all plant sizes. The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 21.

Table 21. COST/BENEFIT ANALYSIS RESULTS FOR HYDRAULIC LOAD FOLLOWING IN PHOSPHORUS REMOVAL VIA LIME ADDITION

Plant Size (mgd)	Annualized Instrument Cost ($)	Chemical Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
1	3,390	9,400	36	(2 x 8210)/(9400 - 3390) = 2.8 years
5	3,390	39,000	8.7	(2 x 8210)/(39,000 - 3390) = 0.46 year
10	3,390	72,000	4.7	(2 x 8210)/(72,000 - 3390) = 0.24 year
50	6,780	245,000	2.8	(4 x 8210)/(245,000 - 6780) = 0.14 year
100	13,560	430,000	3.1	(8 x 8210)/(430,000 - 13,560) = 0.16 year

Figure 66 describes the instruments required for a total load following, based on the pH trimback on the hydraulic flow ratio. The corresponding capital and operating costs (63) are shown in Table 22.

The annualized cost of one unit of this control system (64), based on an 8% interest rate for components with 5 years lifespan and on 6% for all others as shown in Table 22 is $3734.

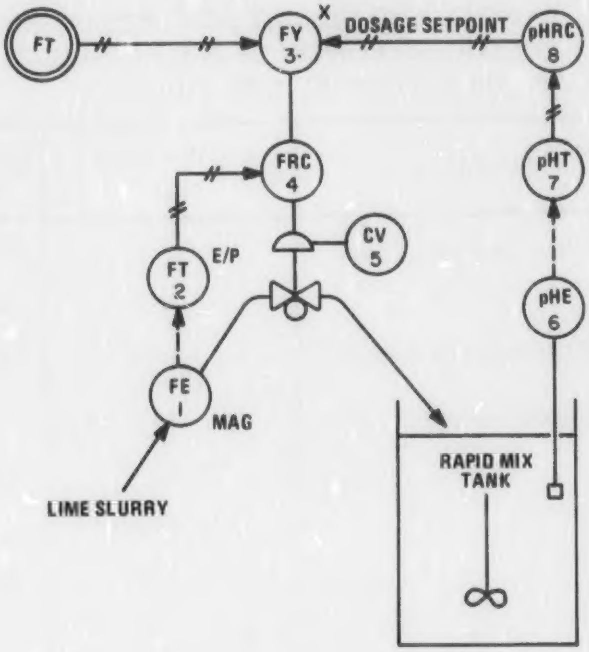

FT
FY 3
X
DOSAGE SETPOINT
pHRC 8
FRC 4
pHT 7
E/P
CV 5
FT 2
E/P
pHE 6
FE 1
MAG
LIME SLURRY
RAPID MIX TANK

Figure 66. Total load following.

One of these systems will be used before each primary clarifier and, therefore, the annualized total cost as a function of plant capacity is:

1 mgd	$ 7,468
10 mgd	$ 7,468
50 mgd	$14,936
100 mgd	$29,872

Considering these yearly costs and the resulting yearly savings (given in Table 19), this control system can also be justified for all plant sizes except the 1 mgd plant. The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 23.

Table 22. CAPITAL AND OPERATING COSTS OF TOTAL LOAD FOLLOWING CONTROLS FOR PHOSPHORUS PRECIPITATION BY LIME ADDITION

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period		
pHE-6	510	50
pHT-7 (with ultrasonic cleaning)	1,000	10
	1,510	
10 years depreciation period		
FT-2	1,200	10
FY-3	500	10
FRC-4	760	16
CV-5	720	12
pHRC-8	820	16
	4,000	
15 years depreciation period		
FE-1	1,400	12
Installation materials	500	
Control panel section	600	30
Engineering and design	2,400	166 hours/year
Installation and startup labor	3,000	
	7,900	Yearly cost of replacement parts is estimated as $340
Total installed cost per loop	13,410	
Total annual operating cost per loop		$2,000

Table 23. COST/BENEFIT ANALYSIS RESULTS FOR TOTAL LOAD FOLLOWING IN PHOSPHORUS REMOVAL VIA LIME ADDITION

Plant Size (mgd)	Annualized Instrument Cost ($)	Chemical Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
1	7,468	13,700	55	(2 x 13,410)/(13,700 - 7468) = 4.3 years
5	7,468	59,400	12.5	(2 x 13,410)/(59,400 - 7468) = 0.58 year
10	7,468	108,500	6.9	(2 x 13,410)/(108,500 - 7468) = 0.27 year
50	14,936	410,000	3.7	(4 x 13,410)/(410,000 - 14,936) = 0.14 year
100	29,872	718,000	4.2	(8 x 13,410)/(718,000 - 29,872) = 0.16 year

Phosphate Precipitation Through Alum Addition

The cost of additional instruments needed for hydraulic and total load following is compared here to the economic benefits of reduced alum usage and to the cost reduction of chemical sludge disposal.

When alum is charged at a fixed rate corresponding to the maximum flow rate and to the maximum pollutant concentration expected in a single day, this mode of operation results in the maximum expense for alum makeup. Sludge disposal may generate additional cost benefits but no economic analysis was performed. The yearly total cost (Table 16) is $24,500 for a million gallons treated per day.

The potential savings, resulting from the application of the hydraulic load following and total load following control strategies, is listed in Table 24. The multipliers used are from Table 15.

Figure 67 describes the instruments required for hydraulic load following. The corresponding capital and operating costs (63) are tabulated in Table 25. The annualized cost of one unit, using 6% interest, is $1695 for the useful lines shown in Table 25.

Table 24. BENEFITS OF AUTOMATION

Plant Size (mgd)	Savings Through Hydraulic Load Following ($/year)	Savings Through Total Load Following ($/year)
1	(0.46) (24,500) = 11,300	(0.67) (24,500) = 16,400
5	(0.38) (122,500) = 46,600	(0.58) (122,500) = 71,000
10	(0.35) (245,000) = 85,500	(0.53) (245,000) = 130,000
50	(0.24) (1,225,000) = 294,000	(0.40) (1,225,000) = 490,000
100	(0.21) (2,450,000) = 514,000	(0.35) (2,450,000) = 858,000

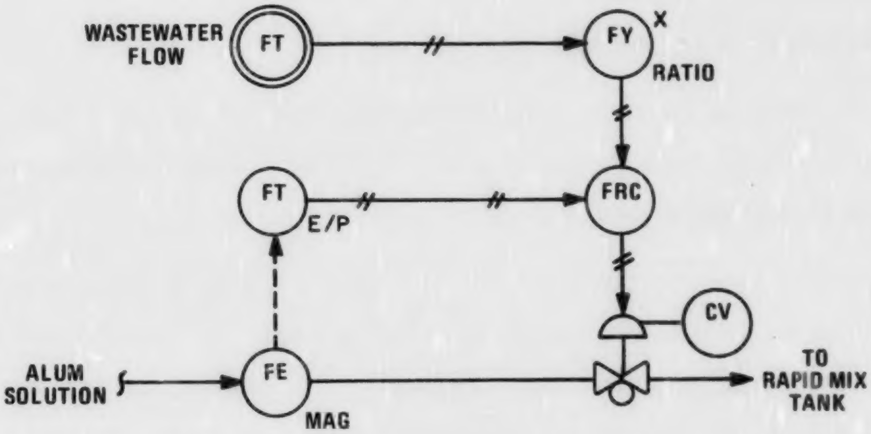


Figure 67. Phosphorus precipitation by alum addition through hydraulic load following control mode.

If one of these loops for each clarifier (Table 17) is used, the annualized cost as a function of plant capacity varies from $3390 for a 1 mgd plant to $13,560 for a 100 mgd plant. The potential yearly savings resulting from this control strategy is shown in Table 24, and the payback periods and other relevant data are furnished in Table 26.

Table 25. CAPITAL AND OPERATING COSTS OF HYDRAULIC LOAD FOLLOWING CONTROLS

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
10 years depreciation period		
FT-2	1200	10
FY-3	380	10
FRC-4	760	16
CV-5	720	12
	3060	
15 years depreciation period		
FE-1	1400	12
Installation materials	300	
Control panel section	350	15
Engineering and design	1400	75 hours/year
Installation and startup labor	1700	
	5150	
Total installed cost per loop	8210	
Total annual operating cost per loop		$750

Figure 68 describes the instruments required for total load following, based on detecting both the flow rate and phosphate concentration in the raw sewage. This feedforward control strategy is limited by the unavailability of a reliable phosphorus analyzer (1). If it is assumed that the phosphorus analyzer costs $7500 and is out of service 20% of the time and that, during this period, only hydraulic load following is practiced, then the capital and operating costs (63) for this system are as shown in Table 27.

The annualized cost of one unit of this control system, based on an 8% interest for components with 5 years lifespan and on 6% for all others, as shown in Table 27, is

Table 26. COST/BENEFIT ANALYSIS RESULTS FOR HYDRAULIC LOAD FOLLOWING IN PHOSPHORUS REMOVAL VIA ALUM ADDITION

Plant Size (mgd)	Annualized Instrument Cost ($)	Chemical Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
1	3,390	11,300	30	(2 x 8210)/(11,300 - 3390) = 2.1 years
5	3,390	46,600	7.3	(2 x 8210)/(46,600 - 3390) = 0.38 year
10	3,390	85,500	4.0	(2 x 8210)/(85,500 - 3390) = 0.20 year
50	6,780	294,000	2.3	(4 x 8210)/(294,000 - 6780) = 0.11 year
100	13,560	514,000	2.6	(8 x 8210)/(514,000 - 13,560) = 0.13 year

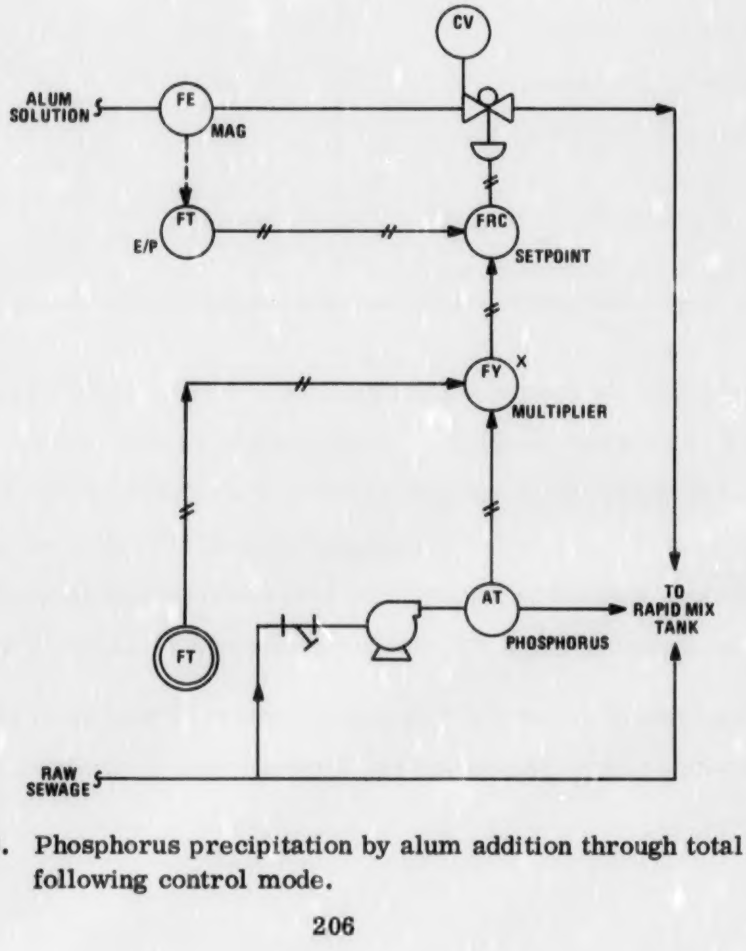

CV
ALUM SOLUTION
FE
MAG
FT
E/P
FRC
SETPOINT
FY
X
MULTIPLIER
AT
PHOSPHORUS
TO RAPID MIX TANK
FT
RAW SEWAGE

Figure 68. Phosphorus precipitation by alum addition through total load following control mode.

Table 27. CAPITAL AND OPERATING COSTS OF TOTAL LOAD FOLLOWING CONTROLS FOR PHOSPHORUS PRECIPITATION WITH ALUM ADDITION

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period		
AT-6 (colorimetric orthophosphate analyzer)[19]	7,500	150
10 years depreciation period		
AT-6 (sampling system, analyzer house, etc.)	2,000	50
FT-2	1,200	10
FY-3	500	10
FRC-4	760	16
CV-5	720	12
	5,180	
15 years depreciation period		
FE-1	1,400	12
Installation materials	500	
Control panel section	600	30
Engineering and design	2,600	290 hours/year
Installation and startup labor	3,200	
	8,300	Yearly cost of replacement parts is estimated as $350
Total installed cost per loop	20,980	
Total annual operating cost per loop		$3,250

$6685 (64). If one of these systems is used in front of each primary clarifier (Table 17), the annualized total cost as a function of plant size is $13,370 for a 1 mgd plant and $53,480 for a 100 mgd plant.

The yearly savings can be obtained from Table 24, based on the assumption that the total load following control strategy will be in operation for 80% of the time, and only hydraulic load following will be practiced during the remaining 20% of the time. Therefore, the projected savings are given in Table 28, and one can conclude that this control strategy can be justified only for plants that are larger than 1 mgd.

Table 28. COST/BENEFIT ANALYSIS RESULTS FOR TOTAL LOAD FOLLOWING IN PHOSPHORUS REMOVAL VIA ALUM ADDITION

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period (No. of Loops) (Installed Loop Cost) / (Yearly Savings—Yearly Costs)
1	13,370	15,360	87.0	(2 x 20,980)/(15,360 - 13,370) = 21 years
5	13,370	66,300	20.0	(2 x 20,980)/(66,300 - 13,370) = 0.79 year
10	13,370	121,100	11.0	(2 x 20,980)/(121,100 - 13,370) = 0.39 year
50	26,740	451,700	6.0	(4 x 20,980)/(451,700 - 26,740) = 0.20 year
100	53,480	790,000	6.8	(8 x 20,980)/(790,000 - 53,480) = 0.23 year

Prechlorination for Odor Control

In this cost/benefit analysis, the potential savings in chlorine consumption resulting from automation is compared to the cost of the required instrumentation.

The main purpose of prechlorination is the destruction of such odor-causing compounds as hydrogen sulfide. This can be accomplished by adding oxidizing compounds that will selectively remove sulfur (ferrous chloride or chlorine). For the purposes of this cost/benefit analysis, it will be assumed that prechlorination will be practiced at a dose of 5 ppm Cl_2 (Table 16). In installations where the total chlorine demand is high or is unpredictably variable, the use of an $FeCl_2$ additive may be a better choice, because less chemical is required for the selective removal of sulfur and because overchlorination—which can slow down subsequent biological treatment—is less likely to occur.

When the chlorinator is set manually for charging chlorine at a fixed rate that corresponds to the product of maximum wastewater flow rate and maximum chlorine demand in the daily cycle, it results in the highest consumption of Cl_2. The yearly total cost for various plant sizes (Table 17) is as follows:

1 mgd	\$ 1,825
5 mgd	\$ 9,125
10 mgd	\$ 18,250
50 mgd	\$ 91,250
100 mgd	\$182,500

The potential total savings, resulting from the application of hydraulic load following and total load following control strategies, is given in Table 29 (the multipliers used are from Table 15.)

Table 29. POTENTIAL BENEFITS OF AUTOMATION

Plant Size (mgd)	Savings Through Hydraulic Load Following (\$/year)	Savings Through Total Load Following (\$/year)
1	(0.46) (1,825) = 840	(0.67) (1,825) = 1,220
5	(0.38) (9,125) = 3,460	(0.58) (9,125) = 5,300
10	(0.35) (18,250) = 6,380	(0.53) (18,250) = 9,650
50	(0.24) (91,250) = 21,900	(0.40) (91,250) = 36,500
100	(0.21) (182,500) = 38,400	(0.35) (182,500) = 64,000

Figure 69 describes the added instrumentation required for hydraulic load following. As explained previously, it is assumed that the chlorinator and the wastewater flow signal exist and are freely available to all control strategies. The capital and maintenance costs (63) for each of the prechlorination control systems are shown in Table 30. The annualized cost of a unit, using a 6% interest rate, is $226 (64) for a useful life of 10 and 15 years, as shown in Table 30. On the basis of using one of these loops for each clarifier (Table 17), the annualized cost as a function of plant capacity is as follows:

1 mgd	$ 452
5 mgd	$ 452
10 mgd	$ 452
50 mgd	$ 904
100 mgd	$1,808

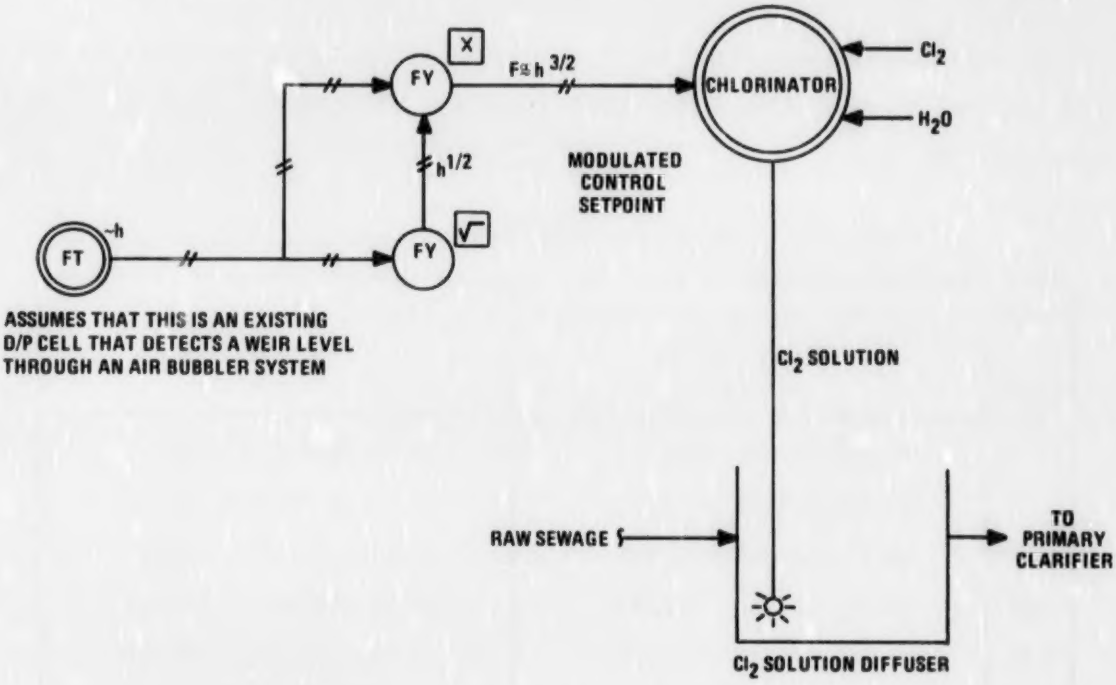

X
FY
F ≅ h 3/2
CHLORINATOR
Cl2
H2O
h1/2
MODULATED CONTROL SETPOINT
~h
FT
FY
√
ASSUMES THAT THIS IS AN EXISTING D/P CELL THAT DETECTS A WEIR LEVEL THROUGH AN AIR BUBBLER SYSTEM
Cl2 SOLUTION
RAW SEWAGE
TO PRIMARY CLARIFIER
Cl2 SOLUTION DIFFUSER

Figure 69. Hydraulic load following controls for prechlorination.

Table 30. CAPITAL AND OPERATING COSTS OF HYDRAULIC LOAD FOLLOWING CONTROLS

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
10 years depreciation period		
FY-1	320	10
FY-2	200	
	520	
15 years depreciation period		
Installation materials	100	
Engineering and design	180	
Installation and startup labor	250	
	530	
Total installed cost per loop	1050	
Total annual operating cost per loop		$100

These costs are lower than the projected yearly savings (Table 29) for all plant sizes. The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 31, together with the projected payback period.

Total load following control is accomplished by modulating the Cl_2 feed rate in accordance with both wastewater flow and concentration of residual chlorine. Figure 70 describes a control system in which the chlorinator is paced by the raw sewage flow signal to give an approximate Cl_2 feed rate, and then is feedback trimmed by oxidation reduction potential (ORP). The ORP is an acceptable measurement for feedback trimming because it measures the relative state of oxidation or reduction of the sewage. The odor-causing anaerobic bacteria function best under negative ORP

Table 31. COST/BENEFIT ANALYSIS RESULTS FRO HYDRAULIC LOAD FOLLOWING CONTROL OF PRECHLORINATION

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
1	452	840	54.0	(2 x 1050)/(840 - 452) = 5.4 years
5	452	3,460	13.0	(2 x 1050)/(3460 - 452) = 0.7 year
10	452	6,380	7.1	(2 x 1050)/(6380 - 452) = 0.36 year
50	904	21,900	4.1	(4 x 1050)/(21,900 - 904) = 0.20 year
100	1,808	38,400	4.8	(8 x 1050)/(38,400 - 1808) = 0.23 year

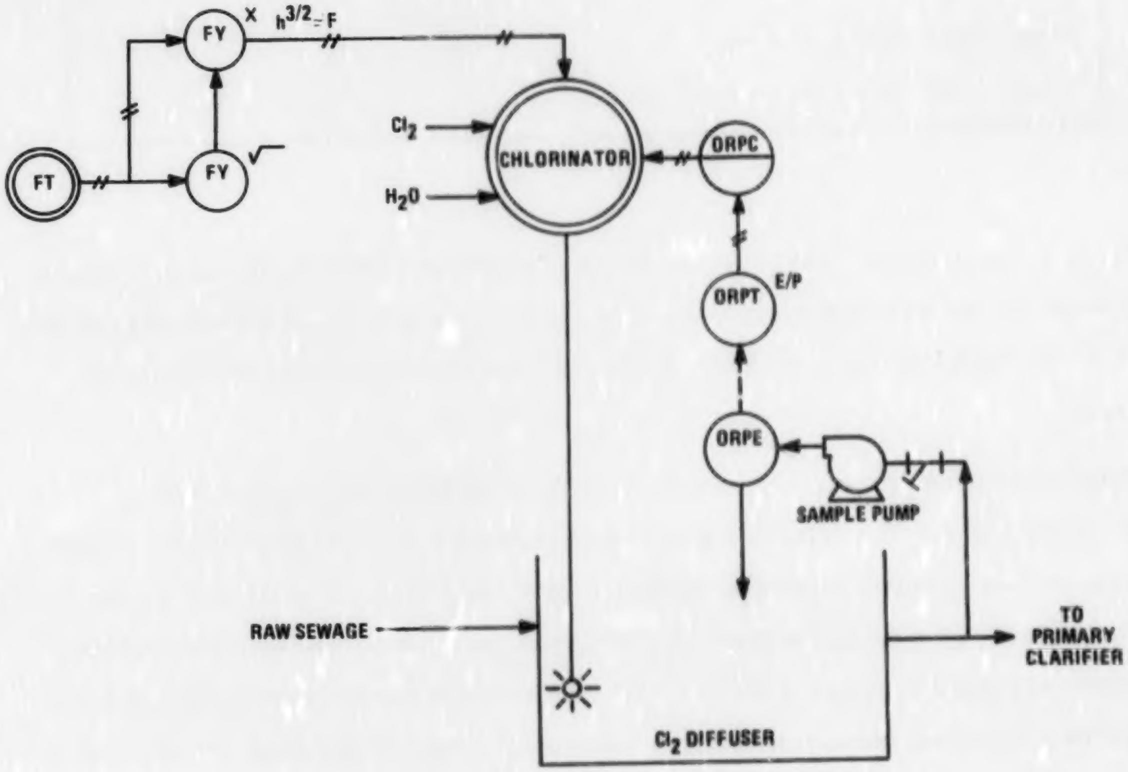

FY
X
$h^{3/2} = F$
FT
FY
√
Cl_2
H_2O
CHLORINATOR
ORPC
ORPT
E/P
ORPE
SAMPLE PUMP
RAW SEWAGE
TO PRIMARY CLARIFIER
Cl_2 DIFFUSER

Figure 70. Total load following controls for prechlorination.

conditions. Another reason for using ORP control is that residual chlorine analyzers cannot be used, because a Cl_2 residual would be wasteful of chlorine and would also retard the biological activity in the downstream aeration tank.

The capital and operating costs for the system (Figure 70), as estimated by Liptak (63) are shown in Table 32. The annualized cost of one unit of this control system, based on an 8% interest rate for components with 5 years lifespan and on 6% for all others, as shown in Table 32, is $2529 (64). If one of these systems is used in front of each primary clarifier (Table 17), the annualized total cost as a function of plant size is $5058 for a 10 mgd or less plant, and $20,232 for a 100 mgd plant.

Table 33 contains the data regarding the overall cost/benefit analysis (the projected yearly savings is based on Table 29). From the cost/benefit analysis, it can be concluded that this control strategy can be economically justified only for plants that are larger than 5 mgd.

Postchlorination for Disinfection

In this cost/benefit analysis, the potential savings in chlorine consumption resulting from automation are compared to the cost of the required instrumentation.

The purpose of disinfection is to prevent the spread of waterborne diseases by eliminating pathogenic organisms. This is guaranteed by adding chlorine in sufficient quantities to provide a residual chlorine concentration of under 1 ppm. Because the chlorine itself is toxic, many areas of the nation require neutralization with SO_2 before the effluent leaves the wastewater-treatment plant in order to protect the receiving surface waters.

When the chlorinator is set manually for charging chlorine at a fixed rate that corresponds to the product of maximum wastewater flow rate and maximum chlorine demand (assumed to be a dose of 15 ppm) in the daily cycle, it results in the highest consumption of Cl_2. The total yearly costs for various plant sizes were presented in Table 16.

Table 32. CAPITAL AND OPERATING COSTS OF TOTAL LOAD FOLLOWING CONTROLS

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period		
ORPE-5	510	50
ORPT-4	1000	10
	1510	
10 years depreciation period		
ORPE-5 (sampling system)	1000	15
ORPRC-3	820	16
FY-1	320	10
FY-2	200	
	2340	
15 years depreciation period		
Installation materials	300	
Control panel section	400	
Engineering and design	1600	10
Installation and startup labor	1900	111 hours/year
	4200	Yearly cost of replacement parts is estimated as $290
Total installed cost per loop	8050	
Total annual operating cost per loop		$1400

Table 33. COST/BENEFIT ANALYSIS RESULTS FOR TOTAL LOAD FOLLOWING CONTROL OF PRECHLORINATION

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period (No. of Loops) (Installed Loop Cost) / (Yearly Savings—Yearly Costs)
1	5,058	1,220	100	Not applicable
5	5,058	5,300	96	Excessive
10	5,058	9,650	52	(2 x 8050)/(9650 - 5058) = 3.5 years
50	10,116	36,500	28	(4 x 8050)/(36,500 - 10,116) = 1.2 years
100	20,232	64,000	32	(8 x 8050)/(64,000 - 20,232) = 1.6 years

The potential total savings resulting from the application of the hydraulic load following and total load following control strategies to chlorination (and dechlorination) are given in Table 34 (the multipliers used are from Table 15).

It is assumed that plants practicing fixed-rate chlorination also practice fixed-rate dechlorination, using liquid SO_2 at $75/ton. The combined economic benefits of load following controls for both a Cl_2 addition and an SO_2 addition are assumed to be 10% higher than for Cl_2 alone.

Table 34. POTENTIAL ECONOMIC BENEFITS OF AUTOMATION OF CHLORINATION

Plant Size (mgd)	Savings Through Hydraulic Load Following ($/year)		Savings Through Total Load Following ($/year)	
	Cl_2 Only	Cl_2 and SO_2	Cl_2 Only	Cl_2 and SO_2
1	(0.46) (5,480) = 2,520	2,770	(0.67) (5,480) = 3,660	4,010
5	(0.38) (27,400) = 10,400	11,400	(0.58) (27,400) = 15,900	17,400
10	(0.35) (54,800) = 19,100	21,100	(0.53) (54,800) = 29,000	31,800
50	(0.24) (274,000) = 66,000	72,700	(0.40) (274,000) = 137,000	150,000
100	(0.21) (548,000) = 115,000	126,000	(0.35) (548,000) = 191,000	209,000

It is assumed that the chlorinator, the wastewater flow signal, and the sulfonator used in the dechlorination step all exist and are freely available to all control strategies. Figure 71 describes the added instrumentation required for the hydraulic load following control strategy. The capital and maintenance costs for the new instruments are given in Table 35.

The annualized cost of one unit, using a 6% interest rate, is \$664 (64) for the useful lives indicated in Table 35. Based on using one of these systems downstream of each secondary clarifier (Table 17), the annualized cost as a function of plant capacity is \$1328 for plants of 10 mgd and less and \$5312 for 100 mgd plants. These costs are lower than the projected yearly savings (Table 34) for all plant sizes. The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 36, together with the projected payback periods.

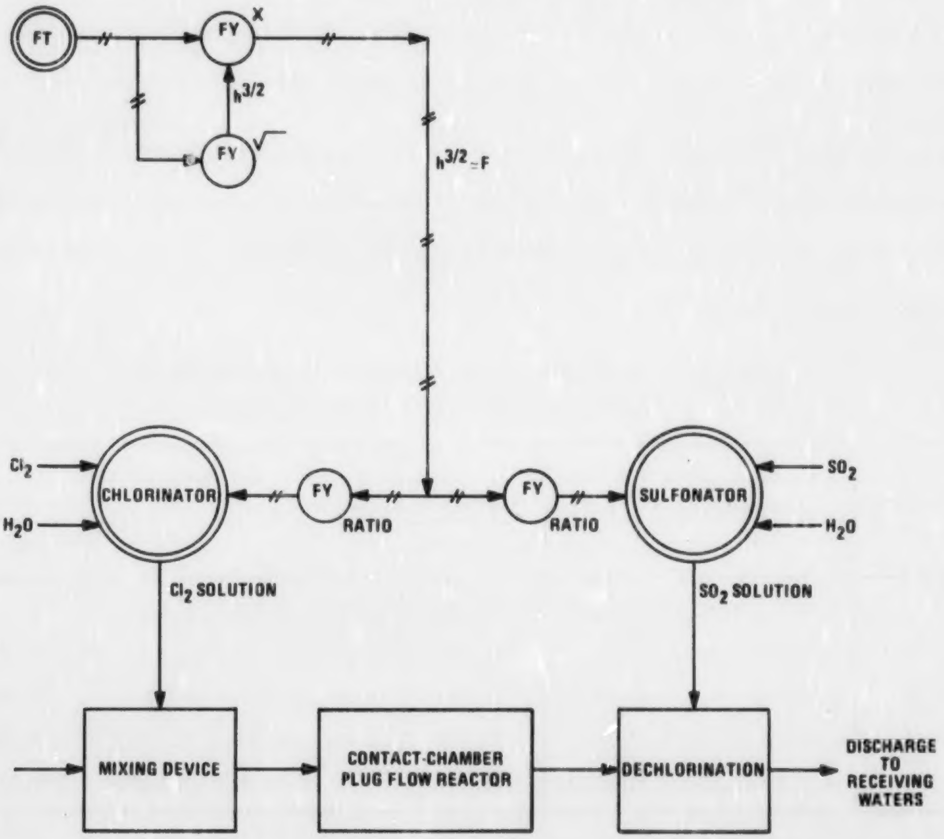

FT
FY
X
$h^{3/2}$
FY
√
$h^{3/2} = F$
Cl_2
H_2O
CHLORINATOR
FY
RATIO
FY
RATIO
SULFONATOR
SO_2
H_2O
Cl_2 SOLUTION
SO_2 SOLUTION
MIXING DEVICE
CONTACT-CHAMBER PLUG FLOW REACTOR
DECHLORINATION
DISCHARGE TO RECEIVING WATERS

Figure 71. Hydraulic load following controls for disinfection and dechlorination.

Table 35. CAPITAL AND OPERATING COSTS OF HYDRAULIC LOAD FOLLOWING CONTROLS

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
10 years depreciation period		
FY-1	320	10
FY-2	200	10
FY-3	320	10
FY-4	320	
	1160	
15 years depreciation period		
Installation materials	200	
Control panel section	300	
Engineering and design	600	
Installation and startup labor	900	
	2000	
Total installed cost per loop	3160	
Total annual operating cost per loop		$300

Total load following is accomplished by modulating both the Cl_2 and SO_2 feed rates in accordance with both wastewater flow and concentration. Figure 72 describes a control system in which both the sulfonator and the chlorinator are paced by the effluent flow rate signal to give an approximate Cl_2 feed rate, and then are trimmed in accordance with residual chlorine concentration. The precontact residual chlorine analyzer (ARC-101) adjusts the chlorinator dosage, and the postcontact chlorine analyzer (ARC-102) trims the setpoint of ARC-101 in a cascade manner. When, due to a sudden upset or misoperation, ARC-102 cannot maintain its setpoint below some preset limit, its output signal is sent through to the sulfonator, activating the SO_2 charging loop (which is normally inactive).

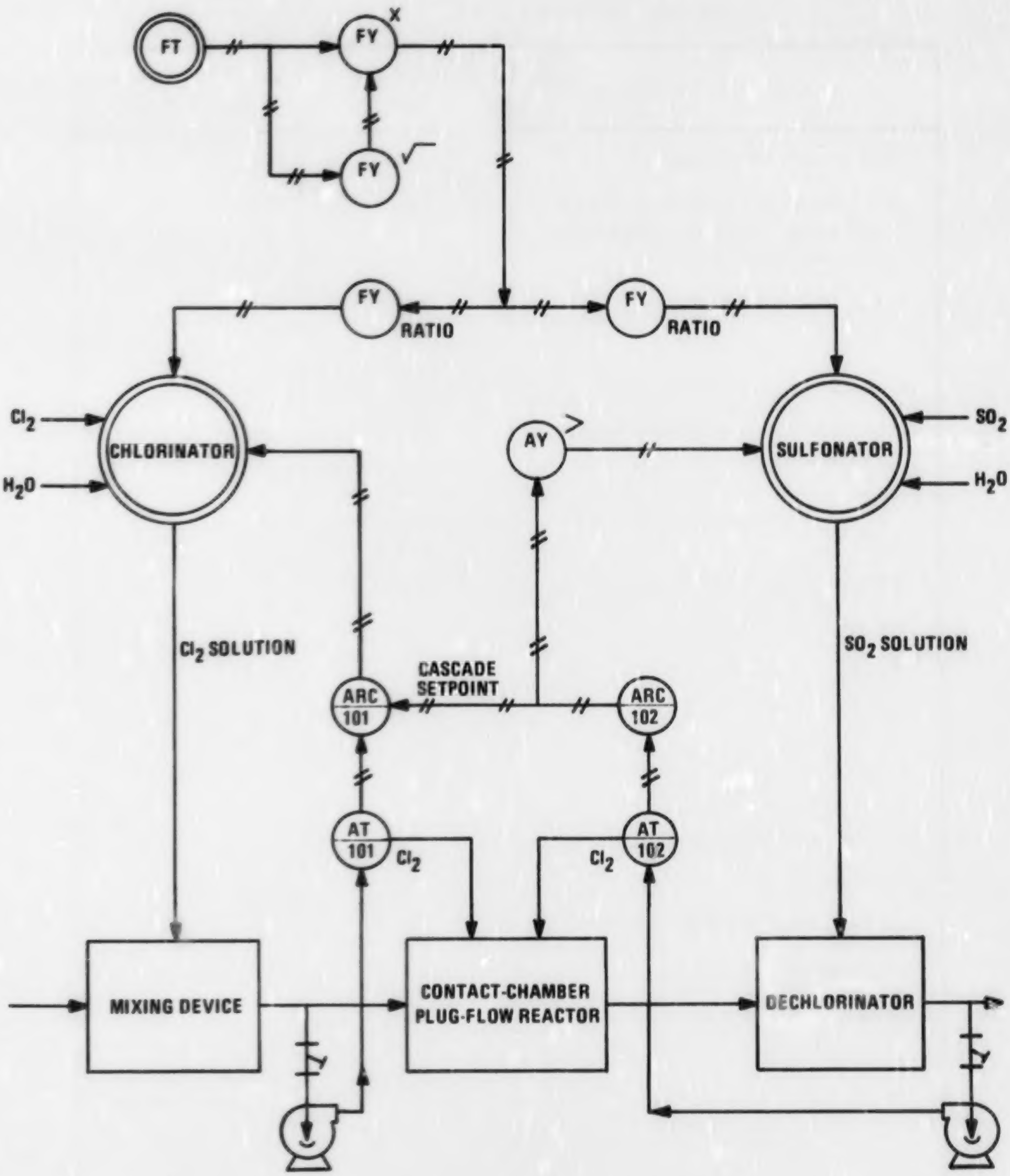
Table 36. COST/BENEFIT ANALYSIS RESULTS FOR HYDRAULIC LOAD FOLLOWING CONTROL OF DISINFECTION AND DECHLORINATION

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period (No. of Loops) (Installed Loop Cost) / (Yearly Savings—Yearly Costs)
1	1,328	2,770	48.0	(2 x 3160)/(2770 - 1328) = 4.4 years
5	1,328	11,400	11.6	(2 x 3160)/(11,400 - 1328) = 0.59 year
10	1,328	21,100	6.3	(2 x 3160)/(21,100 - 1328) = 0.32 year
50	2,656	72,700	3.7	(4 x 3160)/(72,700 - 2656) = 0.18 year
100	5,312	126,000	4.2	(8 x 3160)/(126,000 - 5312) = 0.21 year

The capital and operating costs for the system (shown in Figure 72) were obtained from Liptak (63) and listed in Table 37. The annualized cost of one unit of this control system, based on an 8% interest rate for components with 5 years lifespan and on 6% for all others, as shown in Table 37, is $7466 (64). If one of these systems is used after each secondary clarifier (Table 17), the annualized total cost as a function of plant size is $14,932 for plants less than 10 mgd, $29,864 for 50 mgd plants and $59,728 for 100 mgd plants. Table 38 contains the data regarding the overall cost/benefit analysis. The projected yearly savings are based on Table 34. From the cost/benefit analysis, it can be concluded that, unless the installation of this type of control system is required by law, it can be economically justified only for plants that are larger than 5 mgd.

Aeration

The cost of additional instruments needed for control of DO in activated sludge for hydraulic and total load following control strategies will be compared to the economic benefits of reduced power consumption and increased equipment life.


FT
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SO2
H2O
Cl2 SOLUTION
SO2 SOLUTION
CASCADE
SETPOINT
ARC
101
ARC
102
AT
101
Cl2
AT
102
Cl2
MIXING DEVICE
CONTACT-CHAMBER
PLUG-FLOW REACTOR
DECHLORINATOR

Figure 72. Total load following control for disinfection and dechlorination.

Table 37. CAPITAL AND OPERATING COSTS OF TOTAL LOAD FOLLOWING CONTROLS

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period		
AT-6 (amperometric Cl_2 analyzer with pneumatic output signal and accessories)	3,500	140
AT-6 (same as above)	3,500	140
	7,000	
10 years depreciation period		
AT-6 and AT-7 (sampling system and housing)	3,000	30
ARC-8	820	16
ARC-9	820	16
FY-1	320	
FY-2	200	
FY-3	320	10
FY-4	320	10
AY-5	320	10
	6,120	
15 years depreciation period		
Installation materials	750	
Control panel section	1,000	10
Engineering and design	2,000	382 hours/year
Installation and startup labor	2,400	
	6,150	Yearly cost of replacement parts is estimated as $430
Total installed cost per loop	19,270	
Total annual operating cost per loop		$4,250

Table 38. COST/BENEFIT ANALYSIS RESULTS FOR TOTAL LOAD FOLLOWING

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
1	14,932	4,010	100	Not applicable
5	14,932	17,400	86	Excessive
10	14,932	31,800	47	(2 x 19,270)/(31,800 - 14,932) = 2.3 years
50	29,864	150,000	20	(4 x 19,270)/(150,000 - 29,864) = 0.64 year
100	59,728	209,000	28	(8 x 19,270)/(209,000 - 59,728) = 1.03 years

When the aeration is set at a fixed rate that corresponds to the maximum flow rate and maximum BOD concentration during the daily cycle, it results in a minimum equipment lifespan and in a maximum consumption of power. The total cost of this operation mode equals the sum of the associated operating and capital expenses.

The yearly operating cost can be calculated on the basis that the continuous power consumption (64) for each mgd capacity unit is 20 kW, and that the cost of electricity ranges from 3¢ to 5¢/kWh, as a function of the quantity used. This is summarized in Table 39.

The annualized capital cost of the aeration equipment can be calculated by using the following equation (65):

$$\text{annual capital cost} = (I_{74}/I_{68})\,(1150)\,(P)^{0.81}\left(\frac{i(1+i)^{\eta}}{(1+i)^{\eta}-1}\right) = 140\,(P)^{0.81}$$

where:

I_{74} = present-day Marshall & Stevens cost index of 362

I_{68} = 1968 Marshall & Stevens cost index of 258

Table 39. ANNUALIZED COSTS ASSOCIATED WITH FIXED-RATE AERATION

Plant Size (MGD)	Yearly Operating Costs				Annualized Capital Costs 140 $(P)^{0.81}$ ($/year)	Annualized Total Costs ($/year) (Columns 5 & 6)
	P (Power Used in kW)	Yearly Consumption of Electric Power (10^6 kWh)	Unit Cost of Power Considering Quantity Discount (¢/kWh)	Total Power Cost ($/year)		
1	20	0.175	5.0	8,750	1,650	10,400
5	100	0.875	4.5	39,600	5,740	45,340
10	200	1.75	4.0	70,000	10,250	80,250
50	1,000	8.75	3.5	306,000	37,800	343,800
100	2,000	17.5	3.0	525,000	66,300	591,300
500	10,000	87.5	2.5	2,190,000	244,000	2,434,000

i = assumed interest rate of 6%

η = assumed life expectancy of 20 years

P = fixed maximum electric power requirement in kW (as shown in second column in Table 39)

The annualized operating, capital, and total costs of fixed-rate aeration are shown in Table 39 for the various plant sizes. The potential total savings that result from the application of hydraulic load following, total load following through DO feedback control, and total load following through DO trimmed feedforward TOC control strategies are shown in Table 40. The multipliers are derived from Table 15 and the annualized costs from Table 39.

Figure 73 describes the instruments required for hydraulic load following. The corresponding capital and operating costs taken from Liptak (63) and Table 18 are listed in Table 41. The annualized cost of a single unit of this control system, based

Table 40. POTENTIAL BENEFITS OF AUTOMATION ($/YEAR)

Yearly Savings	Plant Size (mgd)	Benefits of Improved Control		
		Hydraulic Load Following	Total Load Following	
			Feedback (DO)	Feedforward (TOC Trimmed by DO)
		Increased Lifespan (5% = 21 Years)	Increased Lifespan (10% = 22 Years)	Increased Lifespan (15% = 23 Years)
		Reduction in Electric Power Consumption (10%)	Reduction in Electric Power Consumption (25%)	Reduction in Electric Power Consumption (35%)
Savings in operating costs	1	$ 872	$ 2,110	$ 2,900
	5	$ 3,255	$ 8,566	$ 12,000
	10	$ 5,316	$13,780	$ 19,400
	50	$15,312	$45,600	$ 63,700
	100	$23,870	$68,400	$ 95,000
	500	—	—	$298,000
Savings in annualized capital costs	1	$ 9	$ 60	$ 113
	5	$ 56	$ 185	$ 253
	10	$ 96	$ 293	$ 400
	50	$ 240	$ 855	$ 1,300
	100	$ 364	$ 1,290	$ 1,740
	500	—	—	$ 5,500
Total annualized savings	1	$ 881	$ 2,170	$ 3,013
	5	$ 3,311	$ 8,745	$ 12,253
	10	$ 5,412	$14,075	$ 19,800
	50	$15,552	$46,435	$ 65,000
	100	$24,234	$69,690	$ 96,740
	500	—	—	$303,500

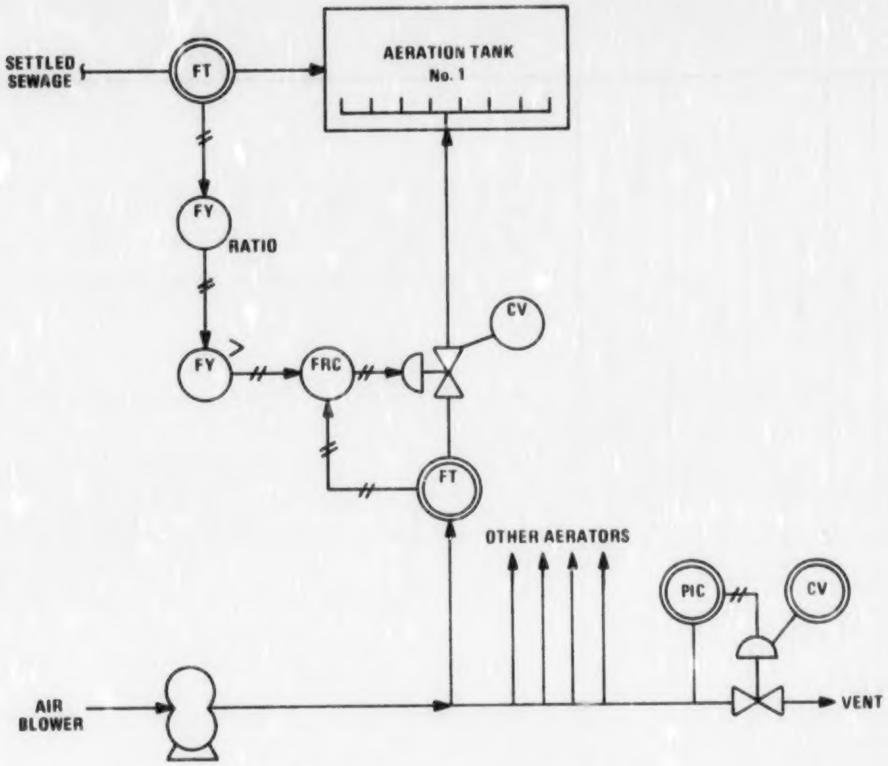


Figure 73. Hydraulic load following control for aeration.

on a 6% interest rate and the data given in Table 41, is, according to Molvar (64), \$1026. It is assumed that one of these systems will be required for each aerator in the plant. Therefore, based on Table 17, the annualized cost as a function of plant capacity is \$2052 for plant sizes of 1 to 10 mgd, increasing to \$8208 for 100 mgd.

Considering the yearly cost of instrumentation listed above and the resulting yearly savings shown in Table 40, this control system can be justified only for plants that are larger than 5 mgd. The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 42.

Figures 74 and 75 describe two total load following control strategies. The first is based on feedback DO control in addition to the hydraulic load following system described in Figure 73, while the second also includes feedforward TOC. The

Table 41. CAPITAL AND OPERATING COSTS OF HYDRAULIC LOAD

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/years)
10 years depreciation period		
FY-1	380	10
FY-2	200	5
FRC-3	760	16
CV-4	600	12
	1940	
15 years depreciation period		
Installation materials	200	10
		53 hours/year
Control panel section	300	
Engineering and design	750	
Installation and startup labor	1000	
	2250	
Total installed cost per loop	4190	
Total annual operating cost per loop		$530

corresponding capital and operating costs obtained from Liptak (63) and Table 18 are given in Table 43. The annualized costs for a single unit of each control system are calculated on the basis of an 8% interest rate for components with 5 years lifespan and 6% for all others, with the data given in Table 43, according to Molvar (64):

annual cost of feedback control = $3110

annual cost of feedforward control = $8290

If one of these systems is installed for each aeration tank (Table 17), the annualized total cost as a function of plant size for both the feedback and feedforward control strategies is as follows:

Table 42. COST/BENEFIT ANALYSIS RESULTS FOR HYDRAULIC LOAD FOLLOWING DO CONTROL

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period (No. of Loops) (Installed Loop Cost) / (Yearly Savings—Yearly Costs)
1	2,052	886	>100	Not applicable
5	2,052	3,300	61	6.5 years
10	2,052	5,400	37	(2 x 4190)/(2494 - 2052) = 2.5 years
50	4,104	15,600	26	(4 x 4190)/(7461 - 4104) = 1.5 years
100	8,208	24,000	34	(8 x 4190)/(11,168 - 8208) = 2.1 years

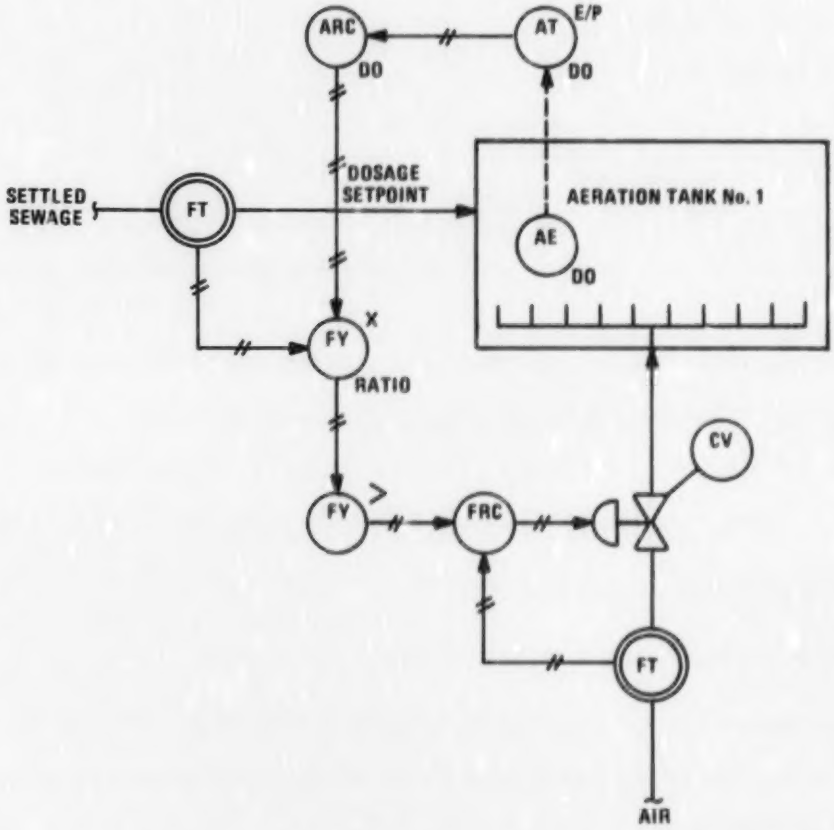

ARC
DO
AT
E/P
DO
DOSAGE
SETPOINT
SETTLED
SEWAGE
FT
AERATION TANK No. 1
AE
DO
FY
X
RATIO
FY
>
FRC
CV
FT
AIR

Figure 74. Total load following based on feedback control.

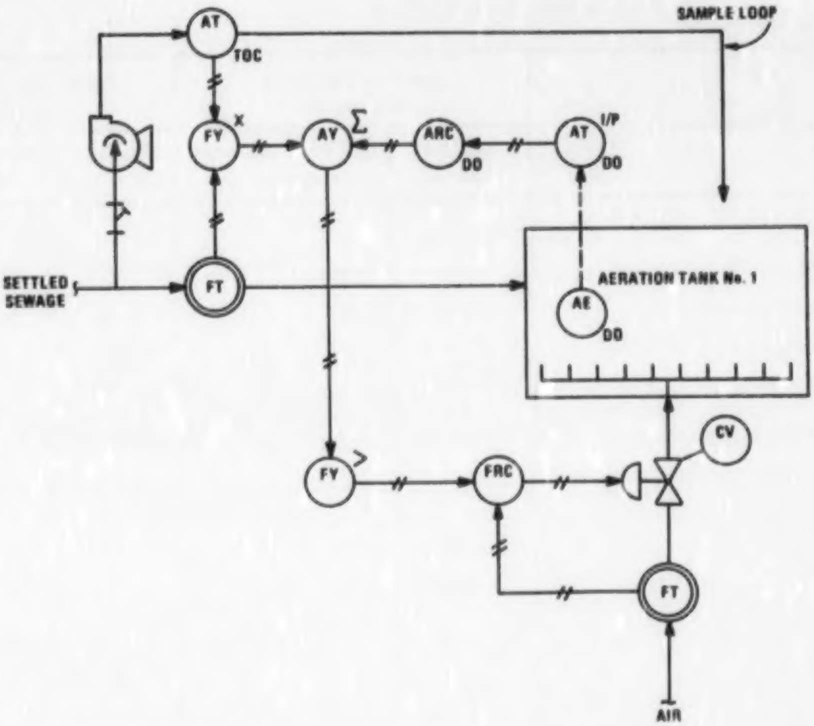

AT
TOC
SAMPLE LOOP
FY
X
AY
Σ
ARC
DO
AT
I/P
DO
SETTLED SEWAGE
FT
AERATION TANK No. 1
AE
DO
FY
>
FRC
CV
FT
AIR

Figure 75. Total load following based on feedforward control.

Size	Feedback	Feedforward
1 mgd	$ 6,220	$ 16,580
5 mgd	$ 6,220	$ 16,580
10 mgd	$ 6,220	$ 16,580
50 mgd	$12,440	$ 33,160
100 mgd	$24,880	$ 66,320
500 mgd	—	$165,800

Based on the assumptions that: 1) the cost of electric power will remain discounted to large users (Table 39), and 2) a complete TOC analyzer is to be dedicated to each

Table 43. CAPITAL AND OPERATING COSTS OF TWO TOTAL LOAD FOLLOWING CONTROLS

Cost Components	Feedback Scheme (Figure 74)		Feedforward Scheme (Figure 75)	
	Capital Cost ($)	Req'd Maintenance (hours/year)	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period				
AE-5 (galvanic DO with cleaner)	800	60	800	60
AT-6 (amplifier-converter)	1,000	10	1,000	10
AT-8 (TOC analyzer)			10,000	150
	1,800		11,800	
10 years depreciation period				
AT-8 (sampling system, housing, etc.)			2,000	25
FY-1	380	10	380	10
FY-2	200	5	200	5
FRC-3	760	16	760	16
CV-4	600	12	600	12
ARC-7	820	16	820	16
AY-9			300	10
	2,760		5,060	
15 years depreciation period				
Installation materials	400		600	
Control panel section	600	15	750	20
Engineering and design	1,600		2,200	
Installation and startup labor	2,100		2,800	
	4,700	144 hours/year	6,350	334 hours/year
		Yearly cost of replacement parts is estimated as $360		Yearly cost of replacement parts is estimated as $660
Total installed cost per loop	9,260		23,210	
Total annual operating cost per loop		1,800		4,000

aeration tank, the potential benefits of feedforward control (Table 40) are less than its annualized cost in all cases except for 50 mgd and 500 mgd plants. If either one or both of these assumptions are changed, the economic justification of feedforward TOC control can become feasible for larger than 5 mgd plants. For these reasons, the feedforward control strategy will no longer be considered in this analysis.

In evaluating the benefits of the feedback control strategy, it is assumed that it will be in service 90% of the time while, for the remaining 10% of the time, only hydraulic load following will be practiced. Based on Table 40, the projected savings are:

1 mgd - (0.1) (691) + (0.9) (2,170) = $ 2,022

5 mgd - (0.1) (3,311) + (0.9) (8,743) = $ 8,200

10 mgd - (0.1) (5,412) + (0.9) (14,075) = $13,500

50 mgd - (0.1) (15,681) + (0.9) (46,455) = $43,460

100 mgd - (0.1) (24,218) + (0.9) (69,690) = $65,140

Considering the data in Table 44, this control strategy can be justified only for plants that are larger than 5 mgd.

Table 44. COST/BENEFIT ANALYSIS FOR TOTAL LOAD FOLLOWING VIA FEEDBACK CONTROL

Plant Size (mgd)	Annualized Instrument Cost ($)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
1	6,220	2,022	>100	Not applicable
5	6,220	8,200	75	9.35 years
10	6,220	13,500	46	(2 x 9260)/(8,680 - 6,220) = 2.5 years
50	12,440	43,365	28	(4 x 9260)/(28,600 - 12,440) = 1.21 years
100	24,880	65,140	38	(8 x 9260)/(43,000 - 24,880) = 1.84 years

Anaerobic Digestion

In this unit operation, the cost of additional instruments is compared to the economic benefits of increased sludge stabilization capacity. Digester capacity is increased by extending its availability and reducing the frequency and duration of digester failures.

The digestion process biologically converts organic materials into methane and carbon dioxide (13). The steady-state material balance between the amount of organic materials fed and the quantity of methane produced by the methane-forming bacteria can be upset by three types of digester failure causes:

- Hydraulic Overloading—Will wash out the methane-forming microbial population at a faster rate than they are produced.
- Organic Overloading—Will result in the accumulation of volatile acids, which tend to lower the pH and inhibit the methane-forming organisms.
- Toxic Overloading—Causes the death of methane-forming organisms.

Several process variables can signal the approach of a digester upset or failure (for instance, a sudden increase in hydraulic or organic load, low pH, or reduced CH_4 production), and several manipulated variables are available to the instrument engineer as "handles on the process," which can be used to restabilize the operation. These include:

- Adjusting the operating temperature
- Modifying the pH by adding a basic reagent
- Changing the sludge feed rate
- Recycling some of the microorganisms that get separated in two-stage digesters.

The costs and benefits of the control strategies based on these considerations will be compared here to the performance of uncontrolled digesters.

The benefits of automated operation will be determined first by establishing the total volume of the digesters and then by estimating the percentage reduction in digestion capacity, which can be obtained by better use through automation. Table 45 provides the basis on which various digester system costs can be estimated (52). As a first approximation, it can be said that single-stage digester systems represent 30%, and two-stage digester systems 40%, of the total capital investment in a wastewater-treatment plant. Table 46 shows the savings in capital costs for various plant sizes, assuming a 10%, 15%, 20%, 25%, and 30% increase in their throughput due to automation.

Approximately 8 Btu/hours are required for each cubic foot of digester volume to maintain its operating temperature at the desired 35° C. This heat requirement is about 25% of the total energy content of the generated digester gas (66).

Table 45. DIGESTER SYSTEM COSTS

Design Adjustment, F_D		Regional Adjustment, F_R	
Type	F_D	Region	F_R
One Stage	1.000	1	1.000
Two Stage	2.358	2	1.061
		3	0.984
		4	1.032
		5	1.073
		6	1.024
		7	1.062
		8	1.024
		9	1.065
		10	1.031

Notes: Installed reactor cost includes purchased cost of reactor, auxiliaries, handling and setting, piping, concrete, steel, instrumentation, electrical, insulation, paint, and indirect costs (prime contractor engineering and construction overhead):

Installed reactor cost (\$)—(Installed base cost) (F_D) (F_R)

Annual maintenance—Will be approximately 2.47% of installed reactor cost.

Operation cost—Reactors require approximately 0.25 operator per shift.

Table 46. CAPITAL COST SAVINGS FROM DIGESTER AUTOMATION

Plant Size (mgd)	Single-Stage Digesters					Two-Stage Digesters				
	10%*	15%*	20%*	25%*	30%*	10%*	15%*	20%*	25%*	30%*
	Capital Cost Savings ($$10^6$)									
1	$0.047	$0.071	$0.094	$0.117	$0.141	$0.064	$0.096	$0.128	$0.160	$0.192
5	0.153	0.230	0.305	0.382	0.460	2.204	0.306	0.408	0.510	0.610
10	0.246	0.370	0.490	0.615	0.738	0.328	0.492	0.656	0.820	0.980
50	0.780	1.170	1.560	1.950	2.340	1.040	1.560	2.080	2.600	3.110
100	1.320	1.980	2.640	3.300	3.950	1.760	2.640	3.520	4.400	5.260
	Annualized Savings** ($/year)									
1	$ 4,850	$ 7,300	$ 9,650	$ 12,500	$ 14,550	$ 6,600	$ 9,900	$ 13,200	$ 16,500	$ 19,800
5	15,700	23,700	31,400	39,300	47,000	21,000	31,500	42,000	52,500	62,700
10	25,300	38,100	50,600	63,200	75,700	33,700	50,600	67,400	84,200	101,000
50	80,500	120,000	161,000	200,000	241,000	107,000	161,000	214,000	268,000	321,000
100	136,000	204,000	272,000	340,000	406,000	181,000	272,000	362,000	453,000	540,000

*Percent increase in throughput (Table 47).

**Based on a 6% interest rate and 15 years depreciation period.

Figure 76 illustrates the temperature control system, which consists of a continuously operating external sludge circulation system, controlled by a temperature cascade loop. The continuous operation is advantageous because it keeps the digester temperature at a fixed value (instead of cycling between limits) and because it contributes to good mixing and agitation.

Table 47 lists the estimated savings for the various types of control strategies.

Table 47. CAPITAL COST SAVINGS THROUGH AUTOMATION (PERCENT OF TOTAL DIGESTER INVESTMENT)

Control Type	Percent Savings	
	Single Stage	Two Stage
Temperature control only (Figure 76)	10	10
pH control only	15	15
Methane control only	15	25
Temperature and pH control (Figure 77)	20	20
Temperature, pH, and methane control (Figures 24 and 25)	25	30

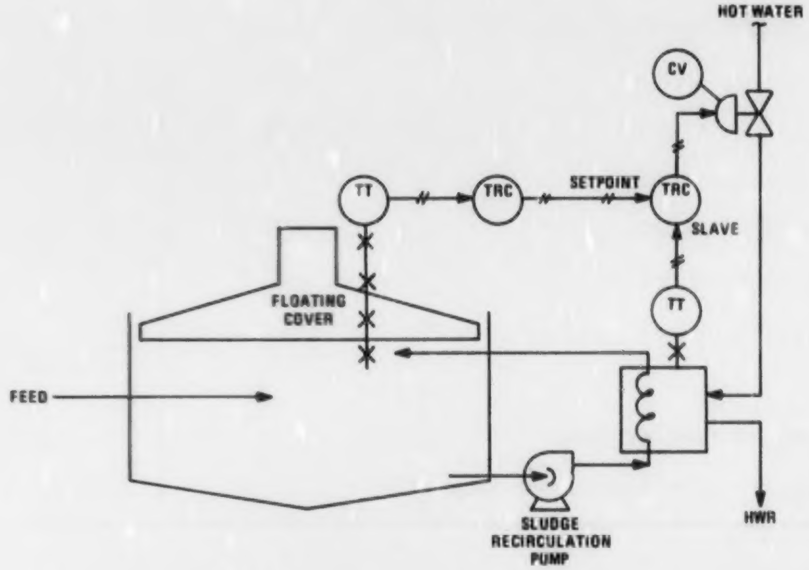


Figure 76. Cascade temperature control of single-stage digester.

The cost of an on-off cycling temperature control system is not substantially lower than the cascade loop described in Figure 76. Capital and maintenance costs are shown in Table 48 (these are the same for both single- and two-stage digestion).

The annualized cost of a single unit for temperature control, using a 6% interest rate, is $1263 based on the information given in Table 48. Using one of these systems for each digester (Table 17), the annualized cost as a function of plant capacity varies from $2526 for a 1 mgd plant to $10,104 for a 100 mgd plant.

Table 48. CAPITAL AND OPERATING COSTS OF TEMPERATURE CONTROL

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
10 years depreciation period		
TRC-2	820	16
TRC-3	760	16
CV-4	500	10
	2080	
15 years depreciation period		
TT-1	400	6
TT-5	350	6
Installation materials	275	
Control panel section	500	6
Engineering and design	900	
Installation and startup labor	1250	—
	3675	60 hours/year
Total installed cost per loop	5755	
Total annual operating cost per loop		$600

These costs are lower than the yearly savings projected for all plant sizes (see the two 10% columns in Table 46). The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 49, together with the projected payback periods.

In addition to good temperature control, it is possible to further improve the automatic control of digesters by continuously monitoring and controlling the pH of the circulated sludge. This will prevent the pH from dropping to the point where the growth of methane-forming microorganisms is inhibited by excessive acidity. The control system necessary to accomplish this is shown in Figure 77. pHE 6 is mounted in an easily isolated bypass line of a well-mixed and continuously flowing sludge sample. Ultrasonic cleaning should increase the mean time between the required preventive-maintenance checks. The nonlinear controller (described previously) should also contribute to a stable control performance.

The capital and maintenance costs shown in Table 50 are the same for both single- and two-stage digestion systems. The annualized cost of a single unit, using an 8% interest rate for devices with 5 years life and 6% for all others, is $3790 using the data given in Table 50. Using one of these systems for each of the digesters (Table 17), the annualized cost as a function of plant capacity varies from $7580 for a 1 mgd plant to $30,320 for 100 mgd plants.

These figures are lower than the yearly savings projected for all plant sizes if the increase in digester capacity realizable is 20% or more (see Table 46). The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 51, together with the projected payback periods.

In addition to the pH and temperature, the quantity of methane gas generated is an important indicator of digester performance. If, at a constant feed rate (fixed hydraulic and organic loading), the methane production drops off, it can be taken as an early signal of a toxic overloading episode. Automatic controls can respond to this

Table 49. COST/BENEFIT SUMMARY OF TEMPERATURE CONTROL OF DIGESTION

Digester Type	Plant Size (mgd)	Annualized Instrument Cost ($/year)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period (No. of Loops) (Installed Loop Cost) (Yearly Savings—Yearly Costs)
Single stage	1	2,526	4,850	52.0	(2 x 5755)/(4850 - 2526) = 5 years
	5	2,526	15,700	16.0	(2 x 5755)/(15,700 - 2526) = 0.87 year
	10	2,526	25,300	10.0	(2 x 5755)/(25,300 - 2526) = 0.5 year
	50	5,052	80,500	6.3	(4 x 5755)/(80,500 - 5052) = 0.31 year
	100	10,104	136,000	7.4	(8 x 5755)/(136,000 - 10,104) = 0.36 year
Two stage	1	2,526	6,600	38.0	(2 x 5755)/(6600 - 2526) = 2.9 years
	5	2,526	21,000	12.0	(2 x 5755)/(21,000 - 2526) = 0.62 year
	10	2,526	33,700	7.5	(2 x 5755)/(33,700 - 2526) = 0.37 year
	50	5,052	107,000	4.7	(4 x 5755)/(107,000 - 5052) = 0.22 year
	100	10,104	181,000	5.6	(8 x 5755)/(181,000 - 10,104) = 0.27 year

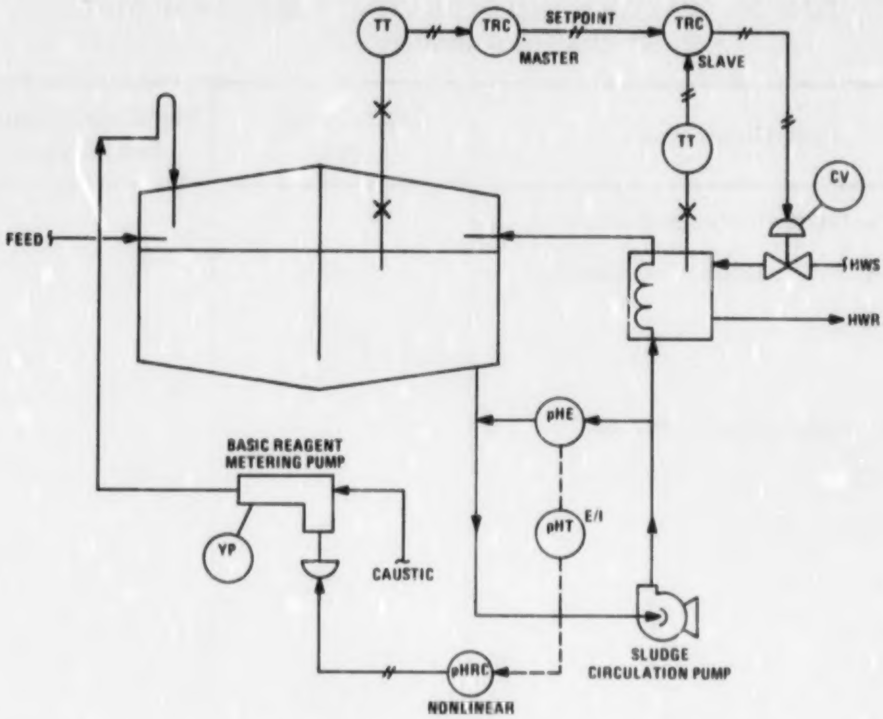

TT
TRC
SETPOINT
TRC
MASTER
SLAVE
TT
CV
FEED
HWS
HWR
pHE
BASIC REAGENT
METERING PUMP
pHT
E/I
YP
CAUSTIC
SLUDGE
CIRCULATION PUMP
pHRC
NONLINEAR

Figure 77. Combined pH and temperature control of single-stage digester.

by increasing the operating temperature in order to increase the growth rate of methane formers or by recycling more methane bacteria from the second-stage separator.

If the dropoff in methane production occurred as a result of increased hydraulic feed rate (resulting in bacteria washout), the automatic instrumentation can reduce or stop the feed, in addition to the corrective actions specified above for toxic overloading.

The instantaneous response to organic overloading is an increase in methane production, which lasts until the resulting low pH starts to inhibit the growth of methane formers. Consequently, organic overloading is best controlled in a feedforward manner by measuring the incoming flow and concentration and keeping its product nearly constant. In the absence of such information, pH can be used as a fairly sensitive feedback control to slow or stop the digester feed.

Table 50. CAPITAL AND OPERATING COSTS OF pH AND TEMPERATURE CONTROL

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period		
pHE-6 (with ultrasonic cleaning)	510	60
pHT-7	1,000	10
	1,510	
10 years depreciation period		
TRC-2	820	16
TRC-3	760	16
CV-4	500	10
pHRC-8	900	16
YP-9	1,000	20
	3,980	
15 years depreciation period		
TT-1	400	6
TT-5	350	6
Installation materials	600	
Control panel section	700	20
Engineering and design	2,400	
Installation and startup labor	3,000	180 hours/year
	7,450	Yearly cost of replacement parts is estimated as $300
Total installed cost per loop	12,940	
Total annual operating cost per loop		$2,100

Table 51. COST/BENEFIT SUMMARY OF pH AND TEMPERATURE CONTROL OF DIGESTION

Digester Type	Plant Size (mgd)	Annualized Instrument Cost ($/year)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period (No. of Loops) (Installed Loop Cost) / (Yearly Savings—Yearly Costs)
Single stage	1	7,580	9,650	79.0	(2 x 12,940)/(9650 - 7580) = 12.0 years
	5	7,580	31,400	24.0	(2 x 12,940)/(31,400 - 7580) = 1.1 years
	10	7,580	50,600	15.0	(2 x 12,940)/(50,600 - 7580) = 0.6 year
	50	15,160	161,000	9.4	(4 x 12,940)/(161,000 - 15,160) = 0.35 year
	100	30,320	272,000	11.0	(8 x 12,940)/(272,000 - 30,320) = 0.43 year
Two stage	1	7,580	13,200	57.0	(2 x 12,940)/(13,200 - 7580) = 4.6 years
	5	7,580	42,000	18.0	(2 x 12,940)/(42,000 - 7580) = 0.75 year
	10	7,580	67,400	11.0	(2 x 12,940)/(67,400 - 7580) = 0.43 year
	50	15,160	214,000	7.0	(4 x 12,940)/(214,000 - 15,160) = 0.26 year
	100	30,320	362,000	8.4	(8 x 12,940)/(362,000 - 30,320) = 0.31 year

The control system that takes all three measurements into consideration is shown in Figure 24 for a single-stage digester, and in Figure 25 for a two-stage digester. In addition to the previously discussed components, this system contains a digester gas composition analyzer (FID or NDIR) and a flow sensor that will yield information on total methane production.

In case methane production drops, the operating temperature can be increased (a single-stage unit), and/or additional methane-forming microorganisms can be recycled (a two-stage unit).

Switches are provided on both pH and methane production to signal low values. These switch actuations can be used automatically to slow or terminate digester feeding or to alarm the operator.

The capital and maintenance costs are listed in Table 52. Two-stage digestion systems require one additional device, the CV-20 valve (shown on Figure 25). The annualized cost for a single unit, using an 8% interest rate for devices with 5 years lifespan and 6% for all others, as shown in Table 52, is $7052. Similarly, the two-stage annualized cost is $7290. Using one of these systems for each digester (Table 17), the annualized cost as a function of plant capacity is given in Table 53.

These figures are lower than the yearly savings projected for all plant sizes except the single-stage digesters in 1 mgd plants (see the 25% and 30% columns in Table 46). The annualized cost of the required instrumentation, expressed as a percentage of the resulting annual savings, is shown in Table 53, together with the projected payback periods.

Table 52. CAPITAL AND OPERATING COSTS OF pH, TEMPERATURE, AND METHANE CONTROL

Cost Components	Single Stage		Two Stage	
	Capital Cost ($)	Req'd Maintenance (hours/year)	Capital Cost ($)	Req'd Maintenance (hours/year)
5 years depreciation period				
pHE-6 (with ultrasonic cleaning)	510	60	510	60
pHT-7	1,000	10	1,000	10
AE-12 (CH_4 analyzer)	4,000	50	4,000	50
AT-13	1,000	10	1,000	10
	6,510		6,510	
10 years depreciation period				
AE-12 (sampling system)	1,000		1,000	
TRC-2	820	16	820	16
TRC-3	760	16	760	16
CV-4	500	10	500	10
pHRC-8	900	16	900	16
YP-9	1,000	20	1,000	20
FY-11	200	5	200	5
FY-14	380	10	380	10
FRC-15	760	16	760	16
FSL-16	75	2	75	2
FAL-17	50	2	50	2
pHSL-18	75	2	75	2
pHAL-19	50	2	50	2
CV-20			1,000	10
	6,570		7,570	
15 years depreciation period				
TT-1	400	6	400	6
TT-5	350	6	350	6
FT-10 (with orifice included)	500	10	500	10
Installation materials	840		840	
Control panel section	900	30	900	30
Engineering and design	3,000		3,000	
Installation and startup labor	4,000		4,000	
	9,990	299 hours/year	9,990	309 hours/year
		Yearly cost of replacement parts is estimated as $510		Yearly cost of replacement parts is estimated as $510
Total installed cost per loop	23,070		24,070	
Total annual operating cost per loop		3,500		3,600

Table 53. COST/BENEFIT SUMMARY OF pH, TEMPERATURE, AND METHANE CONTROL OF DIGESTION

Digester Type	Plant Size (mgd)	Annualized Instrument Cost ($/year)	Savings Resulting From Improved Control ($/year)	Percent of Total Savings Spent on Instrumentation	Payback Period $\frac{\text{(No. of Loops) (Installed Loop Cost)}}{\text{(Yearly Savings—Yearly Costs)}}$
Single stage	1	14,104	12,500	100	Not applicable
	5	14,104	39,300	36	(2 x 23,070)/(39,300 - 14,104) = 1.8 years
	10	14,104	63,200	22	(2 x 23,070)/(63,200 - 14,104) = 0.94 year
	50	28,208	200,000	14	(4 x 23,070)/(200,000 - 28,208) = 0.53 year
	100	56,416	340,000	17	(8 x 23,070)/(340,000 - 56,416) = 0.65 year
Two stage	1	14,580	19,800	74	(2 x 24,070)/(19,800 - 14,580) = 9.2 years
	5	14,580	62,700	23	(2 x 24,070)/(62,700 - 14,580) = 1.0 year
	10	14,580	101,000	14	(2 x 24,070)/(101,000 - 14,580) = 0.56 year
	50	29,160	321,000	9	(4 x 24,070)/(321,000 - 29,160) = 0.33 year
	100	58,320	540,000	11	(8 x 24,070)/(540,000 - 58,320) = 0.40 year

FLOW-AUGMENTING EFFECT OF POLYMERS IN WASTEWATER SYSTEMS

Introduction

The purpose of this analysis is to quantitatively evaluate the potential benefits of flow augmentation through the use of polymers in wastewater transportation systems and then to compare these benefits to the costs of the required polymer addition system.

It has been reported (67) that the addition of polymers to the water flow in open channels will reduce the friction, thereby increasing the velocity of flow. For the purposes of this analysis, the polymer and the concentration to be used are assumed to be Separan AP-30 and 25 ppm, respectively. It is also reported (66) that the flow augmentation effect of this polymer at that concentration is a 25% increase in flow velocity.

Basis of Evaluation

The effect of flow augmentation by adding polymers will be evaluated by analyzing its influence on a typical sewer interceptor. The interceptor can be viewed as a "bottleneck" in a gravity flow collection system. It will be assumed that the distance between the interceptor and the wastewater-treatment plant is 0.75 mile or 4000 feet. Other parameters were assumed or based on information given in "Wastewater Engineering" (67):

■ Pipe diameter	72 inches
■ Manning roughness (N)	0.013
■ Slope	1.8×10^{-4}
■ Design flow velocity	2 fps
■ Design flow	58 cfs

The flood height of the relief well is assumed to be 4 feet. Under storm conditions that result in flooding the well, this extra 4 foot head adds to the energy gradient of the system, resulting in a total maximum flow of 150 cfs.

If the addition of 25 wppm of Separan AP-30 results in a 25% increase in flow, then the total augmented maximum flow is 188 cfs. The gain of this 38 cfs equals the capacity of 42" diameter pipe under flooded conditions. Therefore this cost/benefit analysis will compare the costs of the:

- Construction of a new parallel main, 4000 feet long and 42" in diameter
- Total cost of the polymer addition system.

Assumptions

- The useful life of the 42" parallel main is assumed to be 30 years.
- The interest rate for municipal loans is assumed to be 6%.
- The 42" pipe is assumed to be maintenance free.
- The useful life of the polymer addition system components is in the 10 to 15 years range, as listed specifically in Table 54.
- It is assumed that the polymer addition system will require maintenance as indicated in Table 54.
- It is assumed that flooding conditions exist 1% of the time, or 80 hours a year.
- It is assumed that a suitable building already exists near the interceptor to house the equipment required for the polymer injection.

Cost of 42" Parallel Main

It is assumed that the 42" parallel main can be installed in a relatively unpopulated area with no loss of efficiency due to the breaking of hardtop road surfaces or to the interferences caused by traffic, etc. If these assumptions are not correct, the cost of this system could easily be doubled.

- The material cost at $15/foot of 42" diameter concrete pipe is 4000 x 15 . . . . $ 60,000
- The concrete pipe labor time required at three manhours (MH)/foot of pipe is 12,000 MH. At $18/MH, this gives a total cost of $216,000

Table 54. CAPITAL AND OPERATING COSTS OF POLYMER ADDITION SYSTEM

Cost Components	Capital Cost ($)	Req'd Maintenance (hours/year)
10 years depreciating period		
Feeder (F-1) for 500 lbs/hour capacity	8,000	20
Pump (P-1) for 200 gpm at 10 to 20 feet of head	5,000	10
On-off valve XCV-1	720	12
Five level switches with interlock and time delay	1,500	30
	15,220	
15 years depreciation period		
Bin (T-1), 1000 ft^3 capacity	20,000	
Mixing tank (T-2) provided with a 5 HP agitator (5000 gallon capacity)	40,000	50
	60,000	122 hours/year
Total installed capital cost	75,220	
Total annual operating cost		$1,220

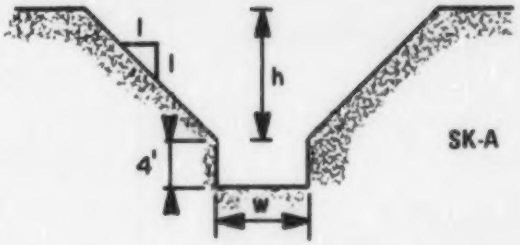
- The excavation and backfill labor cost is estimated in accordance with Figure 78. The required 7200 MH at $18/MH will give a total cost of $130,000
- The cost of sand for backfill is estimated at $3/cubic yard for a total quantity of 10,000 cubic yards . . . . $ 30,000
- Therefore the total installed cost is $436,000

The annualized cost of the 42" parallel main is calculated on the basis of a 30 years lifespan, 6% interest, and zero maintenance cost as:

$$\text{annual cost} = (436,000)\left(\frac{(0.06)\,(1.06)^{30}}{(1.06)^{30}-1}\right) = (436,000)\,(0.0728) = \$31,700$$

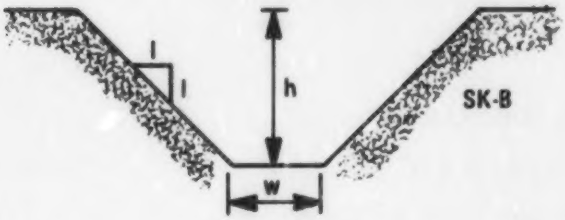
For self-sustaining soil–SK-A:

$$\text{CY/LF} = \frac{4w + (hw) + h^2}{27}$$



For loose soil–SK-B:

$$\text{CY/LF} = \frac{(hw) + h^2}{27}$$



where:

CY = cubic yards of earth
LF = linear feet

- For single lines in a trench, W = outside dimension of line + 1' - 0".
- For multiple runs in a trench, W = sum of outside dimensions of lines + No. of runs x 0' - 6".
- Minimum W = 3' - 0".
- Allow ½ CY concrete for each anchor.
- Show calculation for W and CY/LF below or on a supplementary sheet and identify the cross section calculated by number.

For purposes of this example, let:

$$\frac{10 \times 6 + 10^2}{27} = 6 \text{ CY/FT}$$

Then 4000 LF requires 24,000 CY of excavation and backfill.

Figure 78. Underground pipe excavation, backfill, anchors.

Cost of Polymer Addition System

It is assumed that polymer is added only when flooding conditions occur and that it is added at the fixed rate required to produce a 25 mg/l concentration in a 188 cfs flow ($188 \times 62.4 \times 25/10^6 = 0.3$ lb/second = 1050 lb/hour.

It is also assumed that a suitable building already exists to house the equipment required for the polymer injection.

Assuming that polymer will be used for 80 hours each year and that its cost is \$0.25/lb, the total yearly polymer cost is: Cp = 1050 x 80 x 0.25 = \$21,000/year.

The metering pump (P-1) is sized to deliver 1050 pounds of polymer per hour. This corresponds to about 210 gpm when handling 1% concentration liquid. The polymer storage bin (t-1) is sized to hold a 3 month supply of additive or 20 x 1050 = 21,000 lbs. Assuming a 30 lb/ft^3 polymer density, a 1000 ft^3 bin will be more than adequate. (see Figure 79.) The mix and feed tank (t-2) is sized to hold a 15 minute supply of 1% polymer solution. A 5000 gallon tank is sufficient for this purpose $|(100)\,(1050)/(4)\,(8.3) = 3150|$.

The tapwater charge valve (XCV-1) and the polymer feeder (F-1) are both opened by the low-level switch (LSL-1) and are closed by the high-level switch (LSH-2). XCV-1 is sized to pass 100 gpm (this is equivalent to 50,000 lbs/hour). In order to arrive at a 1% solution concentration, 500 lbs of polymer needs to be added every hour. Consequently, the feeder would be set to deliver approximately 10 lbs or 0.3 ft^3 of polymer/minute. The mixing tank agitator is also started by LSL-1 and is stopped by LSH-2, but only after a 10 minute time delay. LSH-3 starts and stops P-1 while LSHH-4 and LSHH-5 actuate alarm points in the treatment plant's control room.

Table 54 listed the various applicable cost elements (52, 57). These are added to the yearly polymer cost given previously to arrive at the total annualized cost:

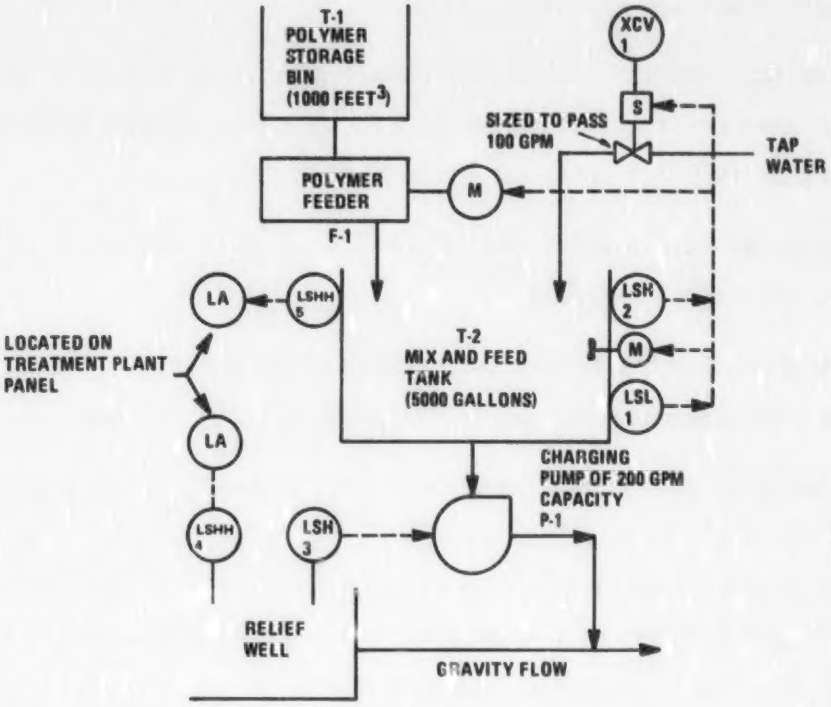

T-1
POLYMER
STORAGE
BIN
(1000 FEET³)
XCV
1
S
SIZED TO PASS
100 GPM
TAP
WATER
POLYMER
FEEDER
F-1
M
LA
LSHH
5
LSH
2
T-2
MIX AND FEED
TANK
(5000 GALLONS)
M
LOCATED ON
TREATMENT PLANT
PANEL
LSL
1
LA
CHARGING
PUMP OF 200 GPM
CAPACITY
P-1
LSHH
4
LSH
3
RELIEF
WELL
GRAVITY FLOW

Figure 79. Polymer addition system for flow augmentation.

$$\text{total annual cost} = 15,220 \left(\frac{(0.06)\ (1.06)^{10}}{(1.06)^{10} - 1} \right) + 60,000 \left(\frac{(0.06)\ (1.06)^{15}}{(1.06)^{15} - 1} \right)$$

$$= \$30,490$$

Conclusions

The annualized cost of installing a 42" parallel main is $31,700 and the annualized cost of a polymer addition in order to achieve the same result is $30,490. On this basis it can be concluded that, if the various assumptions made in this analysis are applicable, the flow-augmenting effect of a polymer addition can be cost competitive with the installation of a new main.

Favorable to the flow augmentation technique is the consideration that the required system can be quickly installed and that its installation would not interfere with traffic or with other aspects of the daily life of a community. The main disadvantage of this approach is the substantial use of polymer, which in itself is a limited resource.

It does appear that the polymer addition strategy, at best, should be considered only as a temporary solution for application wherever the construction of parallel mains would take too long. The polymer charging equipment can be built into a skid so that it will be portable and reusable elsewhere, once it has fulfilled its original function in a particular location.

SECTION IX

MODERN CONTROL SYSTEMS

ADVANCED CONTROL TECHNOLOGY APPLIED TO WATER AND WASTEWATER TREATMENT

Introduction

The two forms of control used almost exclusively in water- and waste-treatment plants are dosage control and simple feedback regulation. Dosage control (also known as pacing, flow proportioning, or ratio control) is used when on-line measurements of the controlled variables are unavailable. An example of dosage control would be the feeding of flocculant in proportion to the flow of the waste stream in order to regulate effluent turbidity. If the quality of the wastewater and the efficiency of the clarifier are reasonably uniform, the effluent turbidity should not change appreciably.

When the controlled variable can be automatically measured, simple feedback control is usually employed. This would be the case wherever the controlled variable is liquid level, pH, residual chlorine, or the like. Although feedback control can provide the precision lacking in dosage control, it also introduces some problems. If the gain of the feedback control loop becomes too high, oscillations will develop; sustained cycling will result in off-specification effluent and can consume chemicals out of proportion to the plant requirements. If the control-loop gain is too low, variations in wastewater flow or composition can cause the controlled variable to deviate substantially from acceptable limits.

The simple feedback controller can be adjusted for only one set of plant conditions so, if the plant or waste characteristics change appreciably, too high or too low a loop gain could result because of nonlinear input/output relationships. Occasionally, the process is particularly difficult to control because the controlled variable is very sensitive to corrective action (as in pH neutralization), or very slow to respond (due to long sampling delays, poor mixing, etc.), or both. In these cases, even if the plant characteristics are constant, moderate variations in wastewater flow and composition can produce an off-specification effluent.

Some industrial treatment plants have an influent equalization basin of several hours capacity which can minimize input variations and thereby facilitate control. But in places where such a capacity is unavailable or the cost is prohibitively high, feedforward control may be the only practical alternative.

This section will discuss the applications of feedforward control to minimize the effects of these influent disturbances, and of adaptive control to compensate for parametric variations within the plant. For a detailed discussion of the theory behind these methods, the reader is referred to work of Shinskey (15, 22). The use of digital computers to implement these and other control methods will then be examined.

Feedforward Control

Feedforward control can be defined as control in which information concerning one or more conditions that can disturb the controlled variable is converted into corrective action to minimize deviations of the controlled variable. Inasmuch as water flow rate is one of the above conditions, dosage control systems that use water flow to set chemical flow are, in fact, feedforward systems. Whether subsequent deviations in the controlled variable (i.e., composition) are continuously recorded or only observed from occasional laboratory analyses is beside the point.

Dosage control systems may take two forms. In the first, the dosage is set but not measured or recorded. In the second, the dosage is calculated from the measurements of two flow rates and, as such, may be controlled and recorded. The calculation of dosage requires a measurement of the manipulated flow, which is not required by the first system.

Figure 80 shows two dosage control systems of the first type. Since the flow of reagent A is the product of stroke and speed, the stroke setting determines the dosage. Reagent B is controlled by a ratio controller—this device multiplies the water flow signal by the setting on the ratio dial to generate the flow setpoint. Since the flow measurements are in the squared form, the dial must have a square-root scale. Figure 81 shows a calculation of the ratio of bandwidth being recorded and controlled; here again, square-root scales are required. Any familiar units may be used, such as mg/l or lb/million gallons.

Occasionally the measurement of the water flow will be separated from the point of the reagent addition by an overflowing vessel. Figure 82 illustrates two such possibilities. If the inflow to the first vessel is measured with chlorine being added to its overflow, the rate of overflow will lag behind the measurement. In this case, a compensating lag is required in the feedforward path. Its time constant should be set to equal the volume change from zero to maximum flow, divided by maximum flow.

If, however, the only measurement available is at the vessel exit, the flow at the point of injection will lead the measurement. Lead-lag compensation is then required. Although a pure lead would be ideal, such a device cannot be made since the ideal function requires infinite gain. The maximum gain available in a real compensator is the lead-to-lag ratio. The lead time constant should be set to match the change in vessel volume from zero to maximum flow divided by maximum flow. The lag time should then be reduced to the lowest value attainable which does not amplify the flow noise excessively.

In actual practice, the compensating time constant varies with the overflowing volume and overflowing rate rather than the maximum values that were selected for setting the compensators. If the overflowing volume varied linearly with the flow, the maximum values would be valid for all flow rates, but such is not the case. It is possible to use

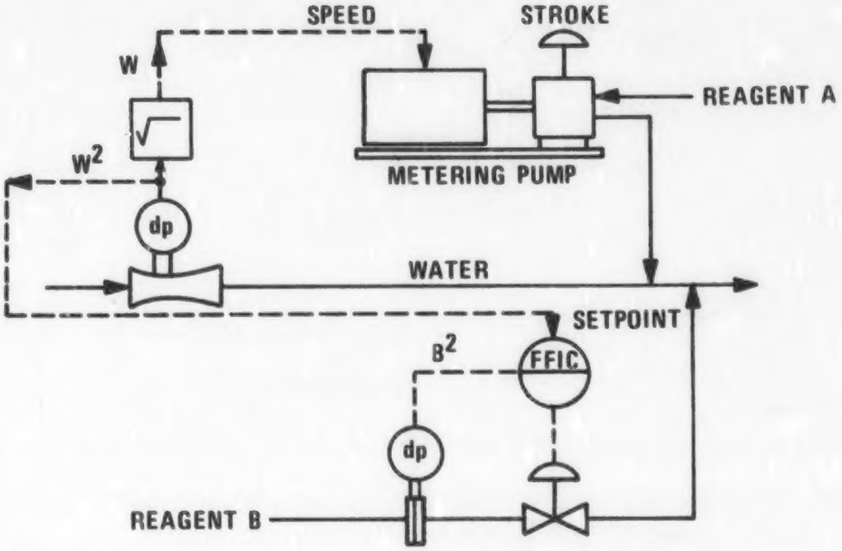

SPEED
STROKE
W
W^2
REAGENT A
METERING PUMP
dp
WATER
SETPOINT
B^2
FFIC
dp
REAGENT B

Figure 80. Stroke of metering pump sets dosage of reagent A, while ratio controller for reagent B has a calibrated ratio or dosage dial.

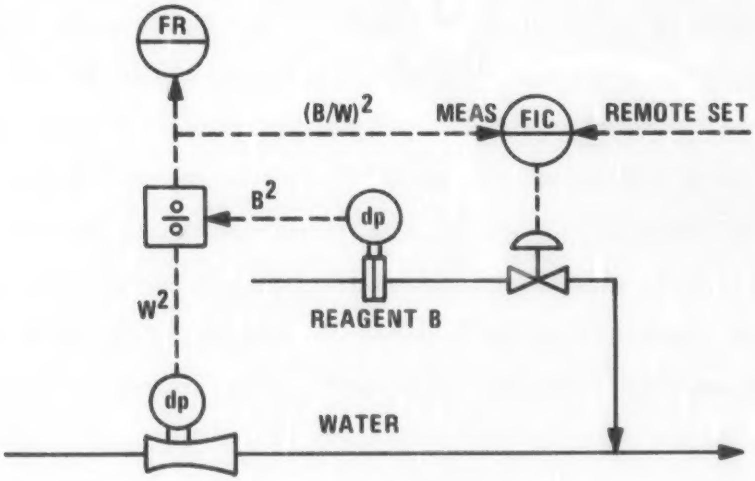

FR
$(B/W)^2$
MEAS
FIC
REMOTE SET
B^2
dp
REAGENT B
W^2
dp
WATER

Figure 81. Dosage calculated by divider may be recorded and is easily set from a remote source such as a digital computer.

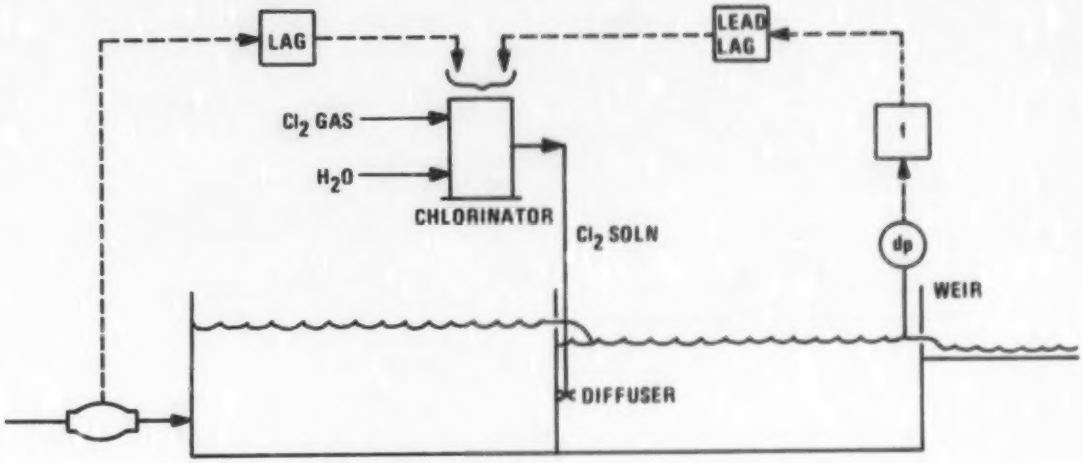


Figure 82. Dynamic compensation is required if actual flow at injection point differs from measured flow in time.

actual measurements of volume and flow to manipulate the settings adaptively but, in most cases, the improvement realizeable is not worth the cost or effort. A number of techniques are available for linearizing the head measurement across the weir for conversion to a usable flow signal (49).

If the particular component requiring treatment in the influent water is measurable, the composition may be used to adjust the dosage. As an example, the alum dosage for phosphorous removal can be based on the amount of orthophosphate in the influent. In most cases, however, the composition measurement is unavailable, nonspecific, insufficient, or too unresponsive to be meaningful. An example of nonspecific measurement is pH. A neutralization process must have base-adjusted to total acid in the influent, but pH indicates only the ionized acid. Consequently, feedforward systems using influent pH measurements have not been as successful as one would like. However, Shinskey (19) reports on the performance of a neutralization facility where effluent specifications could not have been met without feedforward control from the influent pH.

Referring back to the phosphorus removal process, phosphorus loading alone does not determine alum dosage requirements. Alkalinity, other forms of phosphorus, and solids content also have their effect so, even if adequate orthophosphate measurements were available, dosage still could not be adjusted with absolute certainty.

If a sample of the influent must be removed for treatment prior to analysis, the resulting signal will lag behind true process conditions. Thus, a total acidity measurement from a titrated sample would not be useful in most cases for feedforward pH control, because the effluent pH electrodes would be able to respond before a change in influent acidity would be indicated by the titrator.

Figure 83 illustrates a feedforward system using feed composition information. Two precautions are worth noting:

- Dynamic compensation may be required for composition even if not for flow.
- There must be a provision for operating the system without the analyzer in service.

The latter is particularly important since analyzer reliability is not high, and the system cannot function with a zero input.

Both of these requirements can be provided by the controller shown in Figure 83. By connecting its output back to its measurement input, the output will follow the setpoint in the steady state. Furthermore, the dynamic response can be varied by selecting appropriate settings for the controller modes. Finally, upon analyzer failure the controller can be placed in manual, holding its output at the last valid measurement.

Dynamic compensation will differ from that used in the flow signal path because the composition of the entire vessel is affected—not only the overflowing portion. Consequently, the lag will usually be much longer than that applied to the flow signal.

If an on-line analysis of the water quality is available, it will be more usefully applied to the effluent than the influent. For example, it would be impossible to meet effluent pH specifications when using only influent flow and pH for control. A pure feedforward system such as this is not capable of the extreme accuracy required.

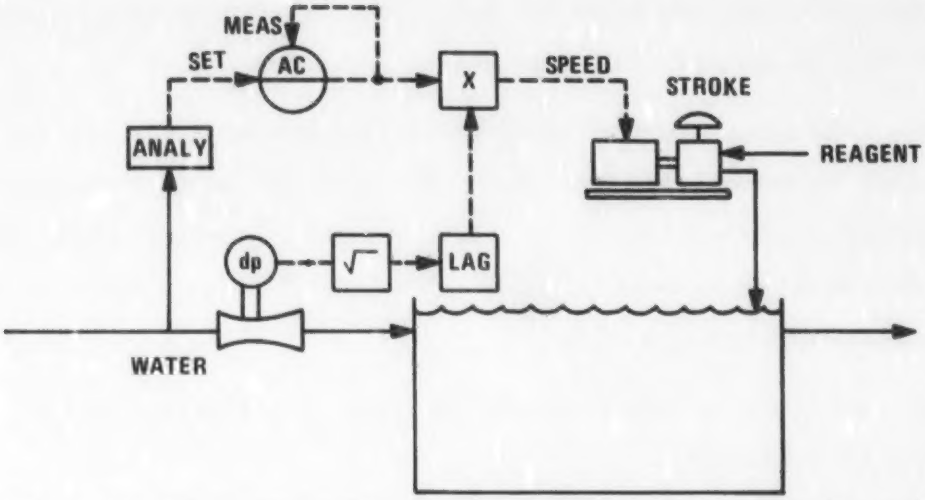


Figure 83. Water composition signal needs dynamic compensation and protection against analyzer failure.

When drawing water from a large source, the flow rate may change rapidly when valves are moved, but influent composition is usually slow to change. Consequently, feedforward control from flow may be necessary while a composition input is not, if an effluent analysis is available for feedback. This logic breaks down to some extent when considering the specialized case of neutralization, however, since influent pH can change rapidly from spills and rinsings. In any case, an effluent quality measurement does reveal the performance of the facility, while the influent composition does not.

Feedback trim from effluent quality is typically applied as shown in Figure 84. If the reagent flow is not transmitted, as with a metering pump or chlorinator, the upper arrangement must be used. If the measurement is available, either configuration is acceptable.

Adaptive Control

Adaptive control can be defined as a control action whereby automatic means are used to change the type or influence (or both) of control parameters in such a way as to improve the performance of the control system. This definition includes a self-adjusting control

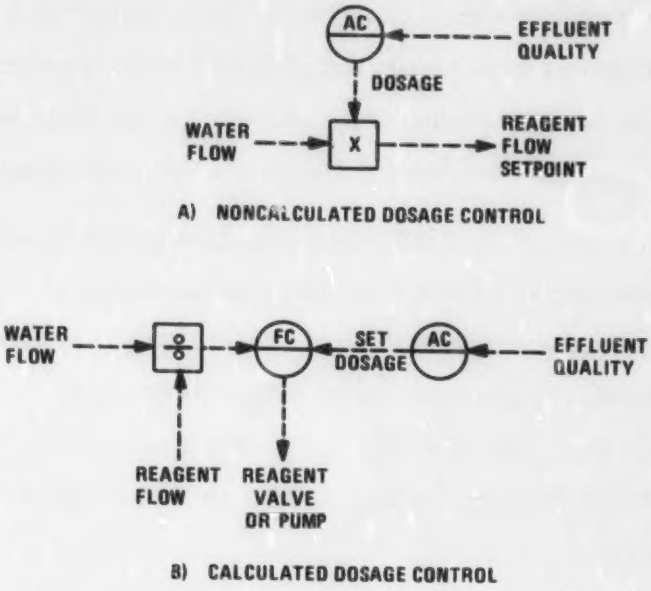

AC
EFFLUENT QUALITY
DOSAGE
WATER FLOW
X
REAGENT FLOW SETPOINT
A) NONCALCULATED DOSAGE CONTROL
WATER FLOW
FC
SET DOSAGE
AC
EFFLUENT QUALITY
REAGENT FLOW
REAGENT VALVE OR PUMP
B) CALCULATED DOSAGE CONTROL

Figure 84. Feedback trim may be accomplished in two ways, depending on whether dosage is calculated.

system, as well as any form of automatic adjustment. The term "adaptive" connotes the control system conforming to the changing characteristics of the process being controlled. By "parameters" is meant the adjustments normally made by hand (e.g., proportional, integral, derivative, lag, gain, etc.).

Adaptive control has not been applied extensively because of the difficulty in adjusting the controller settings by means of a signal. Pneumatic control modes are developed mechanically, with a lever for the proportional band and restrictors for reset and derivative. Automating these adjustments is too clumsy and difficult to be worthwhile, although pneumatic multipliers have been used to adjust loop gain in certain applications.

The mode settings of a typical electronic controller are also introduced mechanically, as the shaft rotation of a logarithmic potentiometer. Shinskey (19) cites an adaptation performed with a nonlinear electronic controller, which is achieved by varying the width of its gap from a remote signal. This has an effect similar to a proportional band adjustment and happens to be ideally suited to pH control situations. It was used in a pH control loop to cancel the effect of the variable gain of equal-percentage valves that were needed to manipulate the reagent flow over a very wide range.

Electronic controllers are now available with modes that can all be adjusted remotely. Although intended primarily to allow a front-of-panel adjustment of a rack-mounted controller, the modes can be set from any 0 to 10 volt signals. Proportional band, reset time, and derivative time are inversely related to the applied voltage. Remote mode adjustment has always been possible with digital computer control. Consequently, most of the research and the few plant installations of adaptive systems have been confined to digital computers.

For many processes, enough information is already available to program the adaption. If the period of a loop varies inversely with flow, for example, proportional, reset, and derivative should all vary inversely with flow. This can be called feedforward adaptation, since the settings are calculated directly from a measurement of the variable that alters the process characteristics. Varying the bandwidth of the nonlinear controller as a function of valve position is another example of feedforward adaptation, since the gain of the loop is known to be affected by the valve position in a predictable manner.

A feedforward control system tries to hold a controlled variable at a desired value by balancing the manipulated variable against a measurement of the load on the process. How successfully it achieves this objective depends on how accurately the required value of the manipulated variable can be calculated, based on the available information on the load and its effect on the process.

Any errors in this calculation result in an imbalance, and hence a deviation, of the controlled variable from its setpoint. This deviation is removed by a feedback controller, which adjusts the feedforward calculation as necessary to compensate for whatever condition may have caused the error. In essence, the feedback controller has adapted the gain of the feedforward loop to match the new process gain.

The top of Figure 84 shows how feedforward and feedback are combined to their mutual benefit in controlling a flowing process. The output of the composition controller adjusts the reagent-to-water ratio to compensate for variations in inlet composition, side reactions, losses, etc. The multiplier also allows the gain of the feedback control loop to be changed directly proportional to water flow, thus achieving a feedforward adaptation of that feedback mode.

In some installations, there is no way of knowing the process gains and measuring those factors that affect it. The most prominent example is the control of pH of plant effluents comprised of a multiplicity of wastes with differing buffer characteristics. The shape of the effluent titration curve may change continuously and randomly. The pH controller can be adjusted only for the conditions that prevail at the time of adjustment; a subsequent reduction in buffering will cause oscillation.

Since oscillations are undesirable, particularly if the controller is alternately adding acidic and basic reagents, the controller must be detuned. The price paid for the absence of oscillations is unresponsive control with all but the least buffered solutions.

In addition to being able to adjust the control modes from a signal, feedback adaptation requires some means of discerning whether an adjustment is needed. If a loop is in the steady state with no deviation, there is no way of knowing its gain. It has to be in a transient or oscillatory condition before its gain and period are revealed.

Most feedback adaptation schemes mimic what an engineer would do in tuning a controller. They introduce step changes in setpoint or controller output, examine the resulting closed-loop response curve, and introduce appropriate changes in settings. But where an engineer may adjust a given controller only once, an adaptive system must function continuously (or at least periodically) if the process characteristics are changing continuously. Needless to say, periodic disturbances intentionally introduced to test the loop are generally undesirable.

An effective adaptive scheme should maintain the loop in an acceptable adjustment by using the normal upsets existing in the process. Even then, two or three oscillation cycles are required before the adjustment is complete, just as when an engineer tunes a controller. In addition, the adaptive device itself needs to be adjusted to avoid instability within its own feedback loop. As desirable as automatic tuning seems to be, it is by no means a panacea. It ought to be reserved for those control loops that can't be stabilized in any other way.

Shinskey (20) gives an example of a case history of a self-adjusting pH control system. By eliminating extended intervals of cycling without sacrificing responsive effluent control, the usage of lime and acid reagents was cut in half. This system is recommended whenever variable buffering is encountered, but especially when both acid and basic reagents are used.

On further work with the system (23), the gap width of a nonlinear controller was adapted to compensate for variations in the titration curve. A discriminator circuit compared the frequency of the pH deviation from the setpoint to an adjustable "crossover" frequency. If the frequency of the deviation was higher, the discriminator developed a positive error signal to expand the gap; if lower, a negative error was generated to close the gap. The crossover frequency was set so that the natural frequency of oscillation would cause the gap to expand. The rate of expansion or contraction was proportional to the pH deviation.

CENTRALIZED CONTROL

Introduction

For the purposes of this discussion, centralized control will be defined as a control system that allows the plant operator to observe and control the plant operation from a remotely located control room. In this control room the main operating parameters are displayed, together with information on emergency or alarm conditions. Provided for the operator's use are: 1) regular analog controllers, which allow him to vary such operating parameters as flows, temperatures, levels, and concentrations, and 2) on-off pushbutton-type remote controls, which allow him to start or stop pumps, open or close valves, etc. The plant's overall status can be quickly observed from the semigraphic panel sections, which show (by steady lights) the valves that are open and the motors that are running. They also indicate (by flashing lights) any operating parameters (e.g., levels, pH, etc.) that are abnormal (see Figures 85 and 86). Visual observation of some unit operations can also be achieved remotely by switching a closed-circuit television display to the corresponding channel. It is also assumed that the analytical laboratory used for grab-sample evaluation (wherever automatic detectors are not yet available) is located in the same building as the control room.

The benefits of central control are many but, in terms of a cost/benefit analysis, labor savings are the most important. In order to provide some quantitative data, the

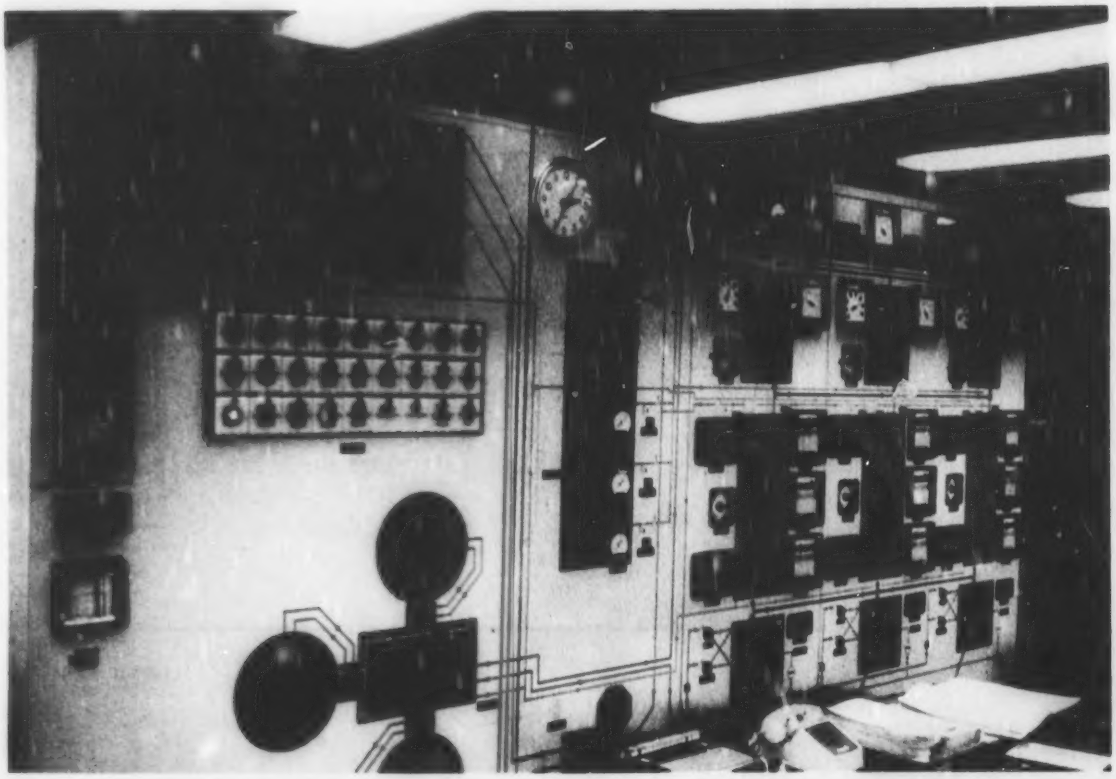


Figure 85. Typical semigraphic display panel in a central control room.

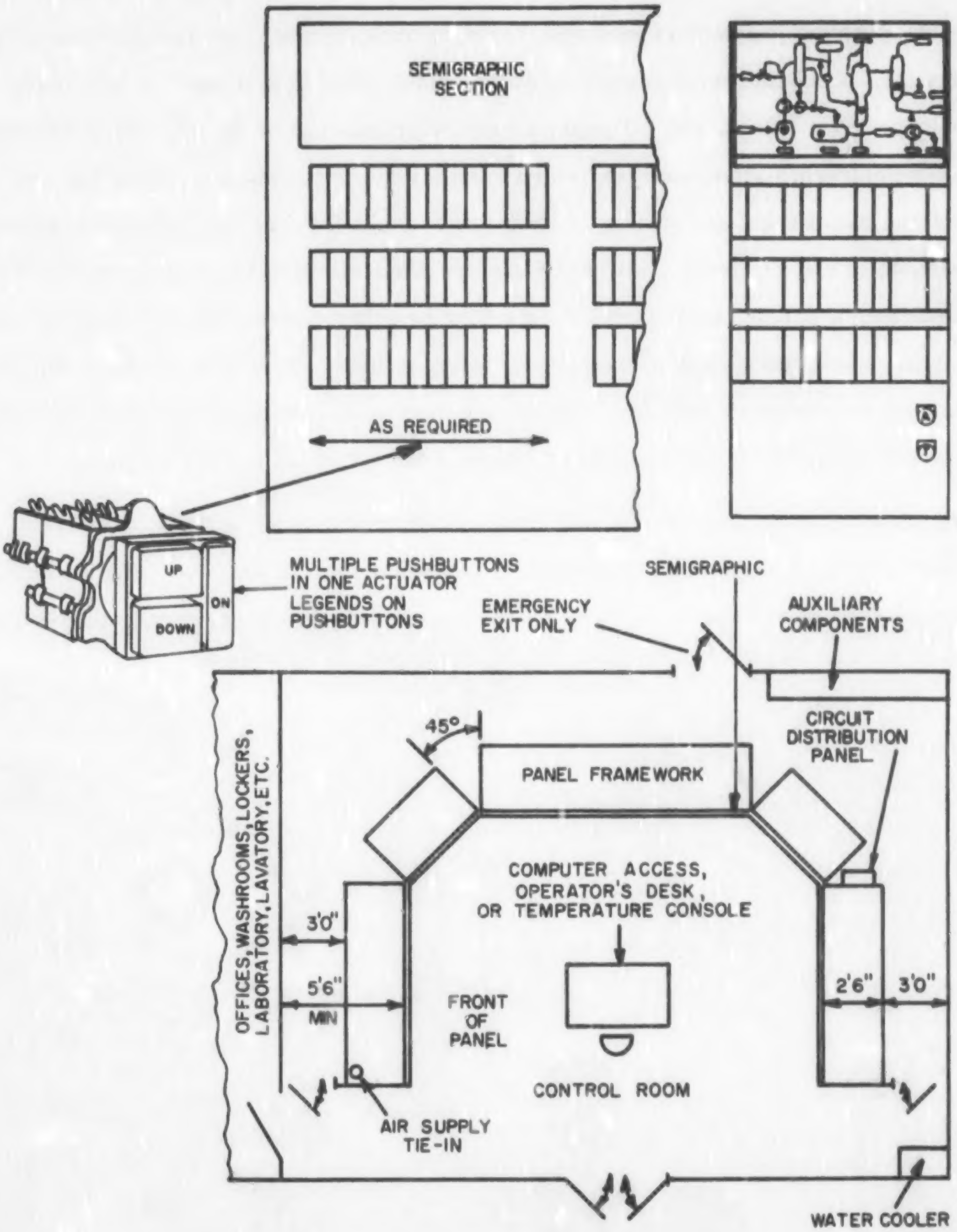

SEMIGRAPHIC SECTION
AS REQUIRED
UP
DOWN
ON
MULTIPLE PUSHBUTTONS IN ONE ACTUATOR LEGENDS ON PUSHBUTTONS
SEMIGRAPHIC
EMERGENCY EXIT ONLY
AUXILIARY COMPONENTS
CIRCUIT DISTRIBUTION PANEL
45°
PANEL FRAMEWORK
OFFICES, WASHROOMS, LOCKERS, LABORATORY, LAVATORY, ETC.
COMPUTER ACCESS, OPERATOR'S DESK, OR TEMPERATURE CONSOLE
3'0"
5'6"
MIN
FRONT OF PANEL
2'6"
3'0"
CONTROL ROOM
AIR SUPPLY TIE-IN
WATER COOLER

Figure 86. Control room layout.

following operating labor estimates illustrate the potential savings of centralized control:

Type of Control System	Operating Costs Spent on Labor
Local control	70%
Central control	50%
Centralized computer control	40%

Table 55 shows the potential total savings as a function of plant size.

The economic justification of centralized control will be developed on the basis of the following assumptions:

- A central building is not available to house the control panel, and its cost at $\$35/\text{foot}^2$ should be included as part of the total cost of centralized control.
- The instrumentation is electronic, and both the signal and 110 volt on-off control leads will be carried in armored conduits.
- The local control system already includes the various transmitters and controllers and, therefore, the cost differential for centralized control should consider only the extra cost of increased transmission distance, not the cost of the instruments.

Table 55. POTENTIAL YEARLY SAVINGS FROM CENTRALIZED COMPUTER CONTROL

Plant Size (mgd)	Centralized Control ($/yr)	Centralized Computer Control ($/yr)
1	12,000	18,000
5	40,000	60,000
10	67,000	100,000
50	220,000	330,000
100	400,000	600,000

- The cost of the nongraphic portion of the central panel is offset by the local panels and support frames needed for the local control strategy.

On the basis of these assumptions, the cost of centralized control will include three main contributing elements: 1) the building cost (given in Table 56), 2) the installed cost of added transmission leads (listed in Table 57), and 3) the cost of a semigraphic section and some miscellaneous alarm, pushbutton, and display devices (outlined in Table 58).

In Table 59, both the savings and costs are summarized and are expressed in both percentage and payback-period terms. Based on this evaluation, centralized control can be justified for all plant sizes.

COMPUTER CONTROL

Introduction

The multipliers and dividers, leads and lags, etc. (shown in Figures 80 to 84), are all computing functions. These functions may be performed (along with feedback control) either with dedicated analog instruments or by a digital computer. This section will discuss the implementation of advanced control strategies by a digital computer.

Since a digital computer represents a substantial investment, it cannot be easily justified if called on to perform only a few tasks. This makes justification difficult for a waste-treatment facility requiring control over only a few variables.

Furthermore, there is little incentive to produce a better quality effluent since it ordinarily has no inherent value. However, if the treated water is to be reused, then the added precision and flexibility of the digital computer may be warranted. This would certainly be the case for a large-scale water-treatment plant, spread over many acres.

Table 56. CONTROL BUILDING COST*

Plant Size (mgd)	Control Panel Length (ft)	Control Room Size (ft x ft)	Control Bldg Area (ft^2)	Total Cost ($)	Annual Cost ($)
1	10	15 x 25	375	13,100	1,020
5	25	20 x 40	800	28,000	2,200
10	30	30 x 40	1200	42,000	3,300
50	50	40 x 50	2000	70,000	5,500
100	60	40 x 60	2400	84,000	6,600

*A depreciation period of 25 years and an interest rate of 6% are assumed. The annual cost is calculated as:

$$\text{total cost} \left(\frac{(0.06)\,(1.06)^{25}}{(1.06)^{25} - 1} \right) = 0.078 \text{ (total cost)}$$

Table 57. SIGNAL AND CONTROL WIRE TRANSMISSION COSTS*†

Plant Size (mgd)	Average Assumed Transmission Distance (ft)	No. of Wire Pairs		Total Transmission Wiring Requirement (No. x ft)	Total Installed Transmission Cost ($)	Annual Cost ($)
		DC Signal (Transmitters, Controllers)	110 Volt On-Off (Pushbuttons, Alarms, etc.)			
1	150	15	40	55 x 150	1,500	118
5	250	25	60	85 x 250	3,850	300
10	300	40	100	140 x 300	7,600	600
50	500	80	200	280 x 500	25,000	1,950
100	600	160	400	560 x 600	60,000	4,700

*A depreciation period of 25 years and an interest rate of 6% are assumed. The annual cost = 0.078 (total cost).

†The installed cost of a 3/4-inch conduit with eight pairs of No. 14 wire is estimated as $1.50/foot. Each foot of required wire pair, therefore, is estimated as 18¢.

Table 58. SEMIGRAPHIC PANEL AND DISPLAY COSTS*†

Plant Size (mgd)	Length of Semigraphic Section Required (ft)	No. of Lights, Pushbuttons, Alarms, etc.	Total Cost of Telephone and Closed-Circuit TV Network ($)	Total Cost ($)	Annual Cost ($)
1	5	30	2,000	7,900	1,070
5	12	45	3,000	16,350	2,220
10	15	75	5,000	22,250	3,030
50	25	150	7,000	36,500	4,960
100	30	300	10,000	49,000	6,650

*The life of this material is assumed to be 10 years, and the corresponding interest rate is assumed to be 6%. The annual cost = 0.136 (total cost).

†The semigraphic cost is assumed to be $1000/foot, and the installed unit cost of the pushbutton-type devices is assumed to be $30 each.

Table 59. COST/BENEFIT SUMMARY OF CENTRALIZED CONTROL

Plant Size (mgd)	Total Projected Savings ($/yr)	Total Projected Costs ($/yr)	Percent of Savings Used to Cover Expenses	Payback Period [Based on Totals From Tables 56, 57 and 58 (Savings and Costs)]
1	12,000	2,208	19.0	22,500/(12,000 - 2,208) = 2.2 years
5	40,000	4,720	12.0	48,200/(40,000 - 4,720) = 1.4 years
10	67,000	6,930	10.0	71,850/(67,000 - 6,930) = 1.2 years
50	220,000	12,410	6.0	131,500/(220,000 - 12,410) = 0.6 year
100	400,000	17,950	4.5	193,000/(400,000 - 17,950) = 0.5 year

Direct Digital Control

Direct digital control (DDC) may be defined as control action in which control is performed by a digital device which establishes the signal to the final controlling element. Historically, DDC was thought to be the most justifiable function of a process computer, and payout was expected simply by replacing analog controllers. In actual practice, DDC was found to exceed the cost of analog control, due principally to the cost of input/output devices and the control stations necessary for backup.

Whether the computer is capable of better control than analog instruments depends on the requirements of the process being controlled. Variables that respond rapidly to manipulation (such as flow, pressure, and liquid level) can usually be controlled better with analog instruments. The reason for this is that the computer does not control continuously but intermittently—at intervals of 1 second or more. The actual scan interval depends not only on access time, but also on the number of tasks assigned to the computer. Analog controllers typically can respond in 0.1 second.

Backup is also a problem with DDC. A computer failure can mean the loss of control throughout the plant. But if the faster loops are controlled by analog instruments, the operators can control the slower ones manually until service is restored. To avoid a complete loss of control over critical variables when the computer is out of service, backup systems like the one shown in Figure 87 can be used. Residual chlorine control is performed in the computer, but chlorine flow is manipulated to meet the desired dosage through analog instruments. Computer outage will cause the computer/manual (C/M) station to transfer to manual, fixing the dosage at its last set value. At any time the operator may adjust dosage from this station. He may also adjust it from the computer console by placing the primary control function analyzer/computer (AC) to manual. Since the computer console may be located some distance from the analog controls, this provision is important.

When full adaptation, or control over processes with relatively long or variable deadtime is required, the computer is capable of superior performance. It should, therefore,

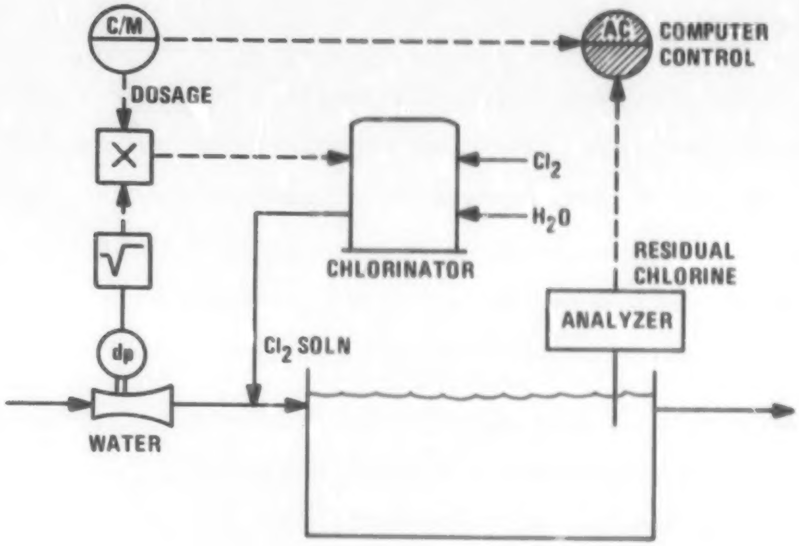


Figure 87. This computer control system fails to provide dosage control.

be saved for these more difficult and higher level operations. In addition to the conventional control modes of proportional, integral, and derivative, and the adaptation thereof, the computer is capable of an endless variety of algorithms. With processes dominated by deadtime, keying the scan interval to the deadtime can result in a substantial improvement over faster scan. Various nonlinear modes (such as error-squared) are also available. In many systems, the engineer is free to compose his own control programs instead of being forced to draw from the conventional ones.

Digital Supervisory Control

Digital supervisory control can be defined as control action in which the control loops operate independently subject to intermittent corrective action; e.g., setpoint changes from an external source. Under the present context, the "external source" is a digital computer, although it also could be simply a remote console supervised by the operator.

Supervisory control systems differ operationally from DDC systems in that the computer manipulates the setpoint of an analog controller. Figure 84 will serve to compare the two. The DDC system with fallback to dosage control applies to the upper schematic in Figure 84—only the C/M station is missing. The lower schematic could be a digital supervisory system if the quality control function (AC) were performed in the computer. This system also has the capability of fallback to dosage control because of the dosage calculation made by the analog divider. A local setting of the dosage is available at the dosage controller or at the computer, without the additional C/M station required in the DDC system. However, the dosage controller is not a conventional analog controller, with the facility for a computer-driven setpoint. It also contains logic to transfer to local set upon computer failure and to signal the computer whenever the controller is placed in local set or manual operation.

Computer Justification

Obviously any in-depth discussions of the capabilities and potentials of computer control would not be feasible within this presentation. However, some general comments on the right reasons and the wrong reasons for installing a computer can be included.

One of the right reasons is the need for solving complex mathematical equations that a human operator could not handle (for example, the maintenance of the optimum food-to-microorganism ratio in activated sludge or digester systems). Another good reason to have a computer is the need for memory and for performing a large number of simple tasks within a short period of time (for example, the regulation of storm-water treatment as a function of several alternate strategies that are stored in the computer memory and automatically selected for use by the computer as a function of the rate, distribution, and duration of the storm). Other tasks that can be given to the computer on a low-priority basis include effluent-monitoring data storage, recordkeeping of all types, inventory control, preventive-maintenance scheduling, and any other administrative and bookkeeping-type function.

One of the wrong reasons for the installation of a computer is to use it as an expensive status symbol. Another is to use it as a replacement for conventional instrumentation without improving the operation.

Wherever a computer control system seems to provide the appropriate solution to a process problem, a choice must then be made between packaged systems for standard applications and custom systems for special requirements.

Packaged systems consist of pre-engineered assemblies of computer hardware and software with suitable sensing actuating, as well as display devices that perform some well-defined and commonly practiced functions. Packaged systems are available for data logging and flow control tasks. Although developed and produced for a fairly wide market, packaged systems do have some flexibility for adapting to process variations and the like. To protect their proprietary interests, many suppliers furnish a "User's Manual" without any program documentation. Packaged systems sell for a clearly stated price and are ideal for users who want the functions that this type of system performs. They are not for users with serious and specialized process control problems, or for those who need a large general-purpose computer system.

For the user who has unique problems that are not addressed by packaged systems or who needs a general-purpose computer system, the solution is to install a custom system. Although a custom system is a unique assembly of equipment and computer programs, its component pieces (especially routine for data acquisition, limit checking, and alarming) are standard. Because of its novel design, a custom system costs more than a packaged system of comparable scope.

The potential yearly savings that can result from computer control has been listed in Table 55. However, the reduction in labor requirements relative to centralized computer control is not very substantial because some of the operating labor requirements that are eliminated by the computer are replaced by an increased need for scheduled preventive maintenance.

Present-day minicomputers are defined more by their price (i.e., less than $20,000) than by their performance. However, prospective users should not be misled by the low cost of the central processor.

Because of the need for special wiring practices and for software development in addition to the hardware costs, even a small digital computer installation is likely to cost $200,000. On the other hand, because many of the algorithms are repetitious and because the size of the computer memory is not the main cost item, even a large computer installation with several CRT displays will not cost more than $800,000. Therefore, the installed cost of central computer control can be taken as the following for various-sized plants:

1 mgd	$200,000
5 mgd	$300,000
10 mgd	$400,000
50 mgd	$700,000
100 mgd	$800,000

Based on 10 years depreciation and 6% interest, the total yearly costs are:

1 mgd	$ 27,200
5 mgd	$ 40,900
10 mgd	$ 54,500
50 mgd	$ 95,000
100 mgd	$109,000

Adding these costs to the annual cost of centralized control (Table 59) gives the total cost of centralized computer control. These costs are compared to the projected savings in Table 60 to show the economic justification and payback periods. It is apparent that computer control cannot be justified unless a plant is larger than 10 mgd.

Although computer control of wastewater-treatment plants is still in its infancy, computer control has become well established in industrial processing. But computer control costs vary widely in both industrial and wastewater processing. The benefits are derived from well-recognized sources, but are not easily predicted. However, as computers become used more frequently in wastewater-treatment projects, an historical background will be built up, and the cost benefits of computer control can then be more quantitatively investigated.

Table 60. COST/BENEFIT SUMMARY OF CENTRALIZED COMPUTER CONTROL

Plant Size (mgd)	Total Projected Savings (Table 55) ($/yr)	Total Projected Costs ($/yr)	Percent of Savings Used to Cover Expenses	Payback Period (yrs)
1	18,000	29,408	>100	Not applicable
5	60,000	45,620	76	Excessive
10	100,000	61,430	62	471,850/(100,000-61,430) = 12 years
50	330,000	107,410	33	831,500/(330,000-107,410) = 3.7 years
100	600,000	126,950	21	993,000/(600,000-126,950) = 2.1 years

SECTION X

INSTRUMENTATION LAYOUTS

INTRODUCTION

The treatment processes discussed in this section and illustrated in Figures 88 to 91 are typical of currently used instrument designs that may not be entirely suitable for any one particular application. These instrument drawings illustrate current practices in the selection and application of instrumentation to wastewater-treatment facilities. The engineering firm of C.E. Maguire, Inc. (located in Waltham, Mass.) was commissioned to produce these instrument designs and the following narrative descriptions.

All instrumentation shown was selected deliberately to perform usefully and reliably with minimum attention. The 1 mgd plants were sparingly instrumented, since it was assumed that the usual minimum team of operators would have little difficulty in keeping the various processes within allowable limits. The 10 mgd plants are much more extensively instrumented, and with the intent of producing useful, reliable information as one of the primary requirements. In all cases, the analyst's role in making periodic and continuing determinations of substrate and solids concentrations and the like is assumed. Completing a typical daily report form is also assumed to be one of the operator's basic duties; consequently, all practical instrumentation to give him the necessary data has been provided.

The existence of a central control room was also assumed, together with the implication that all other field instrumentation would have to perform satisfactorily under corrosive, wet, and often explosive conditions. Recording was separated from the control function, and all signals were assumed to be standard and analog (usually 4 to 20 ma dc or 3 to 15 psi). Separate single-channel recorders are satisfactory for small installations but, wherever there are more than a few recorded or controlled functions, multi-channel recorders are preferable since they show reliably the relative change of variables with time, and with each other. (This is not usually achieved with the customary single-channel recorders.)

RECOMMENDED INSTRUMENTATION

Wherever alarms are shown, highly reliable and industrially proven annunciators are implied. Moreover, their proper utilization is assumed so that the operator is warned when—and only when—a potentially dangerous or costly situation develops.

Level measurement, particularly of wastewater, is most reliably determined by bubble tubes, fitted with constant-flow air supply controllers. Level difference is usually obtained from two bubblers, using a force-balance, non-over-rangeable device of the differential pressure type. The level in digesters and other closed vessels can be reliably measured by force-balance diaphragm-type instruments, but the pressure within the vessel cover must also be considered. Proven optical detectors are available for detecting sludge blanket levels, and sonic devices may also be practical.

Large flows of wastewater are measurable by Parshall flumes, venturis, or magnetic flowmeters. The flumes are simplest and most trouble-free, the venturis must use water-purged connections, and so the magnetic meter is somewhat more practical, although provision must be made for electrode maintenance. Orifices and flow tubes are suitable for air and gas flows, while direct and bypass rotameters should be considered for a flow indication of clean materials.

The radiation density meter remains the best automatic instrument for primary and thickened sludges, but installations should be designed carefully for instrument calibration, servicing, and removal. Density can always be determined rapidly by an operator, using a hydrometer or beam balance and a grab sample; in some cases, this may be the best answer.

The temperature of wastewater is best measured by a platinum resistance bulb or a well-constructed linearized thermistor assembly. The installation of a quality system must be well designed for ruggedness, corrosion-resistance, and serviceability, and can read out to a tenth of a degree with accuracy to ± two-tenths. Filled systems are satisfactory for digesters, while incinerators will have thermocouples supplied with the system.

The weight of sludge on conveyor belts is measurable by many well-proven designs (usually available with the conveyor). Weight may also be of interest in inventory control of chlorine and other materials, but hydraulic systems and more expensive electronic systems are replacing mechanical scales.

pH, dissolved oxygen, and other electrochemical probes are often a problem, especially with raw sewage, but mechanical (and perhaps sonic) cleaning systems are promising. The probes should always be located where they are exposed to a representative section of the process and are easily accessible for inspection and servicing without the need for tools. The probe, probe holder, cable, and junction box or transmitter must be designed to withstand the environment.

Turbidity analyzers, based on the nephelometry of a free surface, are suitable wherever a constant sample stream can be maintained, but they are impractical for raw sewage or sludges and, consequently, are used off-line (i.e., a small representative side stream must be fed to them). Residual chlorine analyzers for both free and total residual are well established and these, too, are used off-line.

Automatic samplers of the type that scoops a small sample at intervals out of a good-sized side stream (e.g., 30 to 50 gpm in a 2" pipe) have been shown to be effective and reliable. However, sample systems employing tubing, peristaltic pumps, an intermittent vacuum, etc., usually do not work on wastewater at all. The "copious side stream" type also adapts itself well, in many cases, to delivering a representative stream through the laboratory for use by technicians, as well as laboratory-type analyzers.

Note also that no sluice gates or large control valves are shown as final control elements. This is because large valves and gates waste energy and because sluice gates in particular must not be used in conventional control loops without adjusting the limited duty factor of the gates.

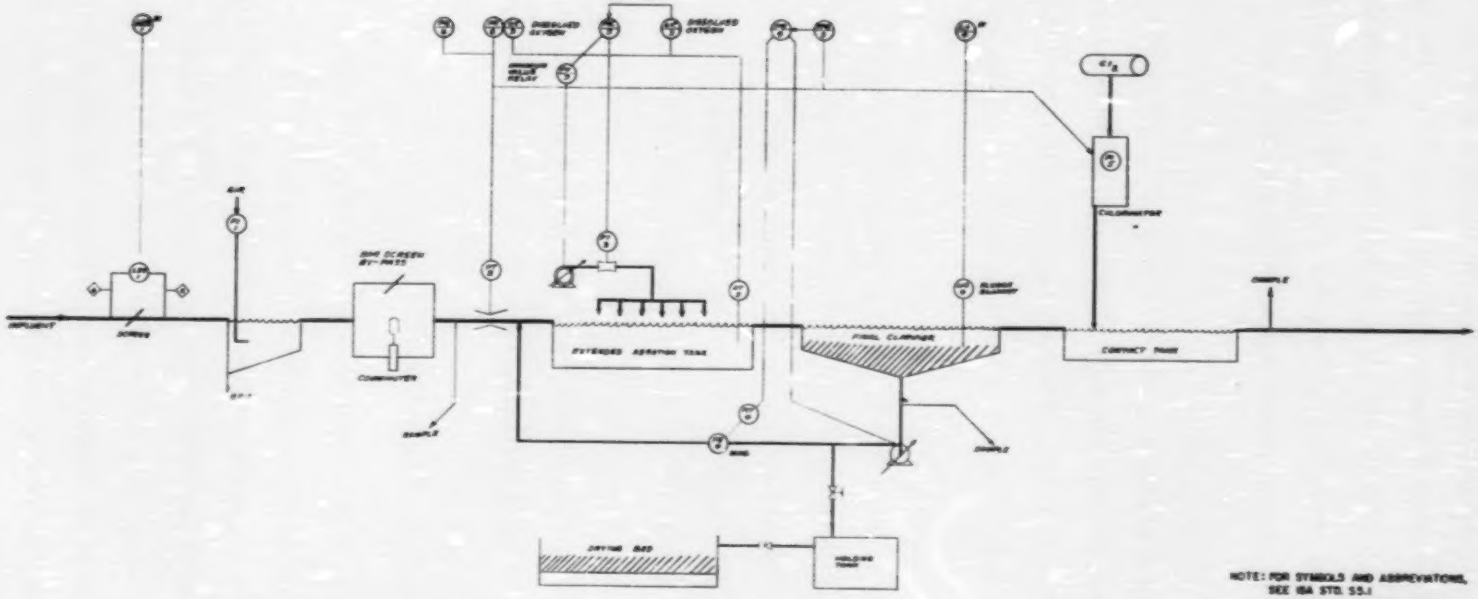
All of the instruments discussed in this section are readily available and can be used with reasonable assurance of a good performance. Their selection and application, however, should be performed, or at least reviewed, by persons who are familar with process rather than laboratory instrumentation, and who are aware of the difference between the two types.

Instrumentation recommendations contained herein are based on the assumption that qualified personnel are available at the applicable site to service and maintain simple analog process-type instrumentation.

ONE MGD ACTIVATED SLUDGE PLANT (Figure 88)

Since the influent to this kind of plant consists of raw sewage, a level alarm (LDA-1) is recommended wherever plugging of the initial screen is possible and of consequence. An airflow meter (FI-1) is recommended wherever air is used to help conserve the air and to make consistent flow adjustments possible. A manual sample point after the comminutor is noted, and total flow is measured with a Parshall flume. The flow is recorded and totalized in a conventional manner. [If the influent flow experiences radical, abrupt changes (as from on-off or high-speed/low-speed pumps), the flume should be moved to wherever the flow changes less radically.] There is no instrumentation associated with the primary clarifier.

The settled influent, mixed with return sludge, is aerated with good mixing and aeration before the dissolved oxygen (DO) probe is reached at AT-3. The DO value is fed back to the DO controller (AIC-3), where a signal is generated to vary the airflow by means of proportional-plus integral controller FRC-3, which maintains the desired value (the setpoint of the DO controller). A minimum value device (FY-3) ensures


INFLUENT
SCREEN
AIR
BAR SCREEN BY-PASS
COMMINUTER
SAMPLE
DISSOLVED OXYGEN
SET POINT
DISSOLVED OXYGEN
MINIMUM VALUE RELAY
EXTENDED AERATION TANK
FINAL CLARIFIER
SLUDGE BLANKET
WAS
SAMPLE
Cl_2
CHLORINATOR
CONTACT TANK
SAMPLE
DRYING BED
HOLDING TANK

NOTE: FOR SYMBOLS AND ABBREVIATIONS, SEE ISA STD. S5.1

CONTROL SYSTEM MAY BE PNEUMATIC OR ELECTRONIC. CONVERTERS ARE NOT SHOWN.

Figure 88. One mgd activated sludge wastewater-treatment plant, instrumentation diagram

that the airflow will provide good mixing and prevent flooding of the air diffusers in spite of any process upset or manual change of AIC-3 or FRC-3. The required airflow (shown on FRC-3) varies with the flow rate and aeration needs of the influent and, by putting the influent flow and DO on the same chart, DO control failures become immediately apparent. The airflow could well be recorded on the same chart, if desired.

Return sludge is pumped at a rate proportional to influent to the plant. FFIC-3 and FIC-4 form a cascaded loop for flow stability, similar to the airflow arrangement with AIC-3 and FRC-3. The flow of return sludge is not recorded (although it can be and it is indicated on the control panel), since functioning of the return sludge system is considered routine.

Good practice recommends a few spare recording points on a control panel, and one of these might be used occasionally for return sludge flow. Any failure of the sludge return system is detected and brought to the operator's attention by the sludge blanket alarm (LA-4). The rest of the system is completely conventional, with the chlorinator paced by the influent flow signal from FT-2 in an open-loop feedforward arrangement. This is usually satisfactory, although the ratio of chlorine flow rate to influent flow rate will not sense a large flow to the holding tank, which could cause a slight but temporary increase in the chlorination ratio (pounds of chlorine: gallons of effluent).

TEN MGD ACTIVATED SLUDGE PLANT (Figure 89)

Influent to this kind of plant is also handled conventionally. Two instruments that can be especially useful are a temperature recorder and a plugged screen alarm. The temperature measurement is intended to show thermal upsets, but the need for TR-1 and LDA-1 depends on conditions peculiar to each site. Flow indicators FI-1A and FI-1B are recommended, since they allow the operator to know how much chlorine or air he is using and thus conserve material and maintain constant flows. The measurement flume and the transmitter (FT-2) provide a signal for recording and totalizing with the addition of a high-flow alarm (FA-2).

Flow equalization is provided by the storage wet well and the variable-speed pumping system. Expected variations in influent rate and the desired degree of smoothing dictate the necessary storage well size and pump flexibility, using conventional control engineering methods. The instrumentation is designed to provide variable-speed floating control with nonlinear forcing at the extremes; in other words, the pumps will change speed only gradually except when the wet well approaches either a full or empty condition, in which case the pumps are driven to a high or low speed in order to keep the wet well level within bounds. The size and shape of the wet well, pumping flexibility, influent variability, and instrumentation all must be considered as a single design.

The operation of the flow equalization system is evident when changes occur. When the equilibrium is stable, the pump output equals the influent rate, and the level in the wet well nears midpoint. An increase in the influent rate is essentially integrated by the wet well, and the pump rate rises in proportion to the level increase in the wet well (caused by an increase in the wet well input over the output). The rate at which the pump output changes in response to the level change is determined by the gain setting of the derivative relay (FY-2B) and that of the controller (LIC-2); this is indicated by the middle slope of the curve on the plot of depth vs outflow in Figure 89. The scaling of the level and flow transmitters, the cross section of the wet well, the pump rangeability, the gain of FY-2B and LIC-2, the integration rate of LIC-2, the characteristics of FY-2A, and the breakpoints in the response curve of LIC-2 all form a complicated relationship that is best designed and tested by means of simulation. The reset and derivative value are particularly sensitive, especially since the wet well and LIC-2 form two integrations in one loop (a potentially unstable situation unless the time constants are quite different).

The low-pass filter (FY-2A) is necessary to exclude noise and other higher frequency disturbances that cannot indicate a true change in the wet well content. The derivative relay (FY-2B) anticipates influent changes and may or may not be necessary, although it can reduce the wet well size requirement; this relay should not be replaced by a

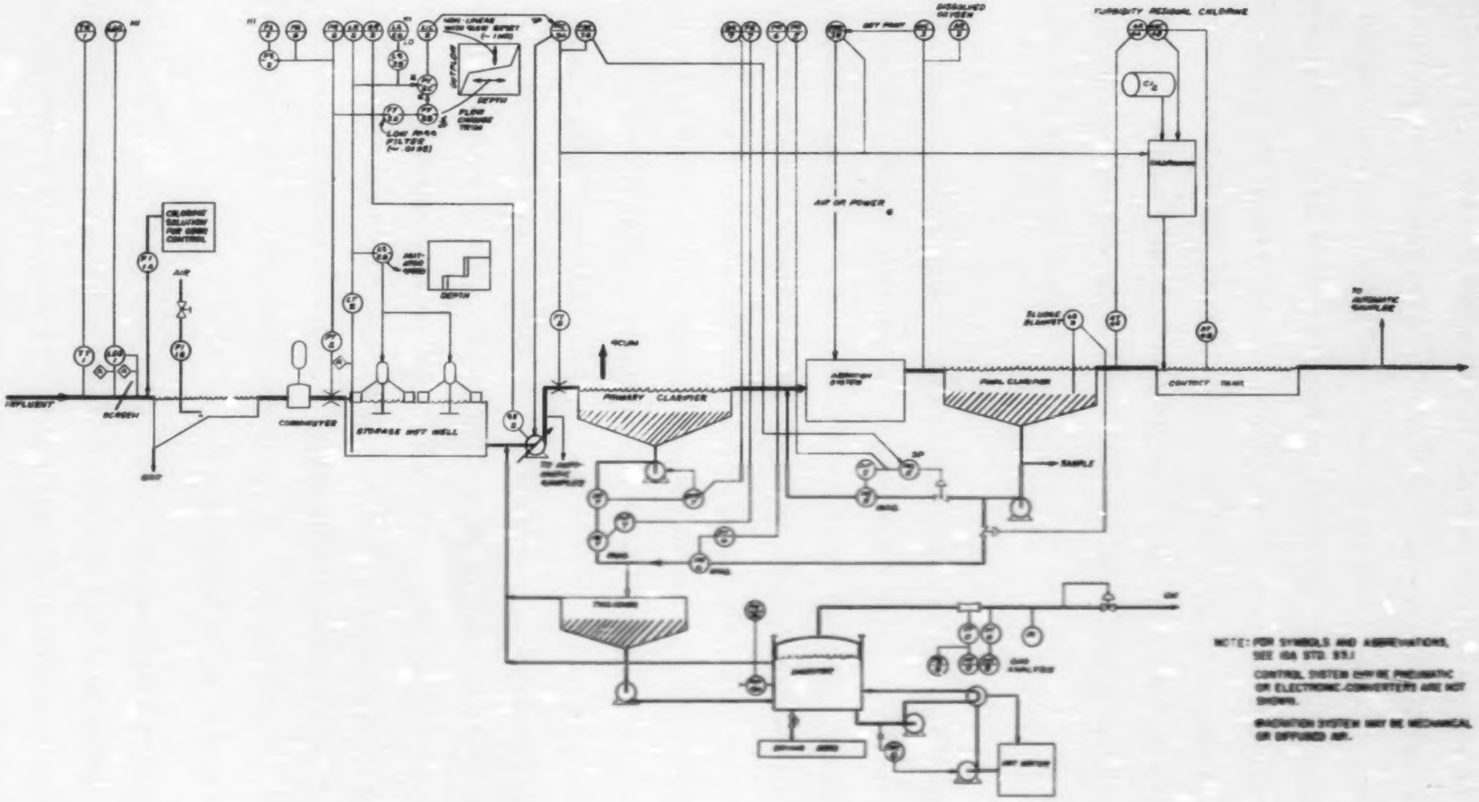

INFLUENT
SCREEN
GRIT
AIR
CHLORINE SOLUTION FOR ODOR CONTROL
COMMINUTER
STORAGE WET WELL
DEPTH
NON-LINEAR WITH WIDE BAND GAP
OUTFLOW
FLOW CHANGE TRIM
LOW PASS FILTER
TO AUTOMATIC SAMPLER
SCUM
PRIMARY CLARIFIER
AIR OR POWER
AERATION SYSTEM
DISSOLVED OXYGEN
FINAL CLARIFIER
SLUDGE BLANKET
SAMPLE
TURBIDITY RESIDUAL CHLORINE
Cl_2
CONTACT TANK
TO AUTOMATIC SAMPLER
THICKENER
DIGESTER
GAS ANALYSIS
NOTE: FOR SYMBOLS AND ABBREVIATIONS, SEE ISA STD. S5.1
CONTROL SYSTEM EITHER PNEUMATIC OR ELECTRONIC-CONVERTERS ARE NOT SHOWN.
AERATION SYSTEM MAY BE MECHANICAL OR DIFFUSED AIR.

Figure 89. Ten mgd activated sludge wastewater-treatment plant, instrumentation diagram

conventional derivative action in a controller (LIC-2) that would be influenced by any setpoint change as well as any input signal change. The performance of the flow equalization process is continuously displayed and then stored in a form usable for future analysis by the three-channel recorder (FR-2/LR-2/SR-2). The flow signal from FT-3 could be substituted for the speed signal, as is shown subsequently in the discussion of the 1 mgd physical/chemical plant.

The level switch arrangement (LS-2B) controls the operation of the aerators/agitators so that they slow down and stop as the wet well level drops, thus preventing excessive bottom scouring and/or the aerators running with too little coverage.

Primary sludge is withdrawn in the conventional manner by using a radiation-type density meter (DE-7) and timers. A sludge blanket level probe could provide equivalent control, but the density meter, when used with FE-7, provides a semiquantitative record of the weight of withdrawn solids. The aeration system is intended to operate similarly to the 1 mgd activated sludge plant discussed previously, with the suggested addition of a feedforward loop (via FFRC-3B) to make the aeration proportional to the flow from the pumps.

Secondary sludge recycle is flow proportioned, as in the previous example, while a sludge blanket probe ensures wasting in the case of a secondary sludge buildup. Moreover, in practice the control valve is usually prevented from closing entirely to ensure continuous wasting. The operation of the final clarifier is monitored by the turbidity recorder (AR-4A), which is useful for indicating trends and upsets but cannot be depended on for a reliable reading of the solids content.

Chlorination is conventional, with the open-loop flow pacing from FT-3 readjusted by closed-loop control from ARC-4B. As usual, the final residual chlorine in the effluent must be determined from samples taken after the chlorine contact tank, since the values from AT-3 are required for control and will be higher than the final values.

The operation of the thickener to concentrate the sludge and feed the digester is not shown; neither are systems to warn of escaping gas and/or control the withdrawal of waste sludge or gas. The single exception to this is a back-pressure valve to ensure that digester pressure is always positive and sufficient to support the digester cover. A conventional method to circulate and heat the digester contents is shown, using a thermostatic device (TIC-8) which controls the hot-water pump.

The digester gas analyzer(s) indicated by AT-8 are useful for sensing changes in digestion, rather than merely the average composition. Suitable analyzers might monitor the methane content (via flame ionization detectors) or simply density, but they work if—and only if—the sample transport and conditioning system is well designed.

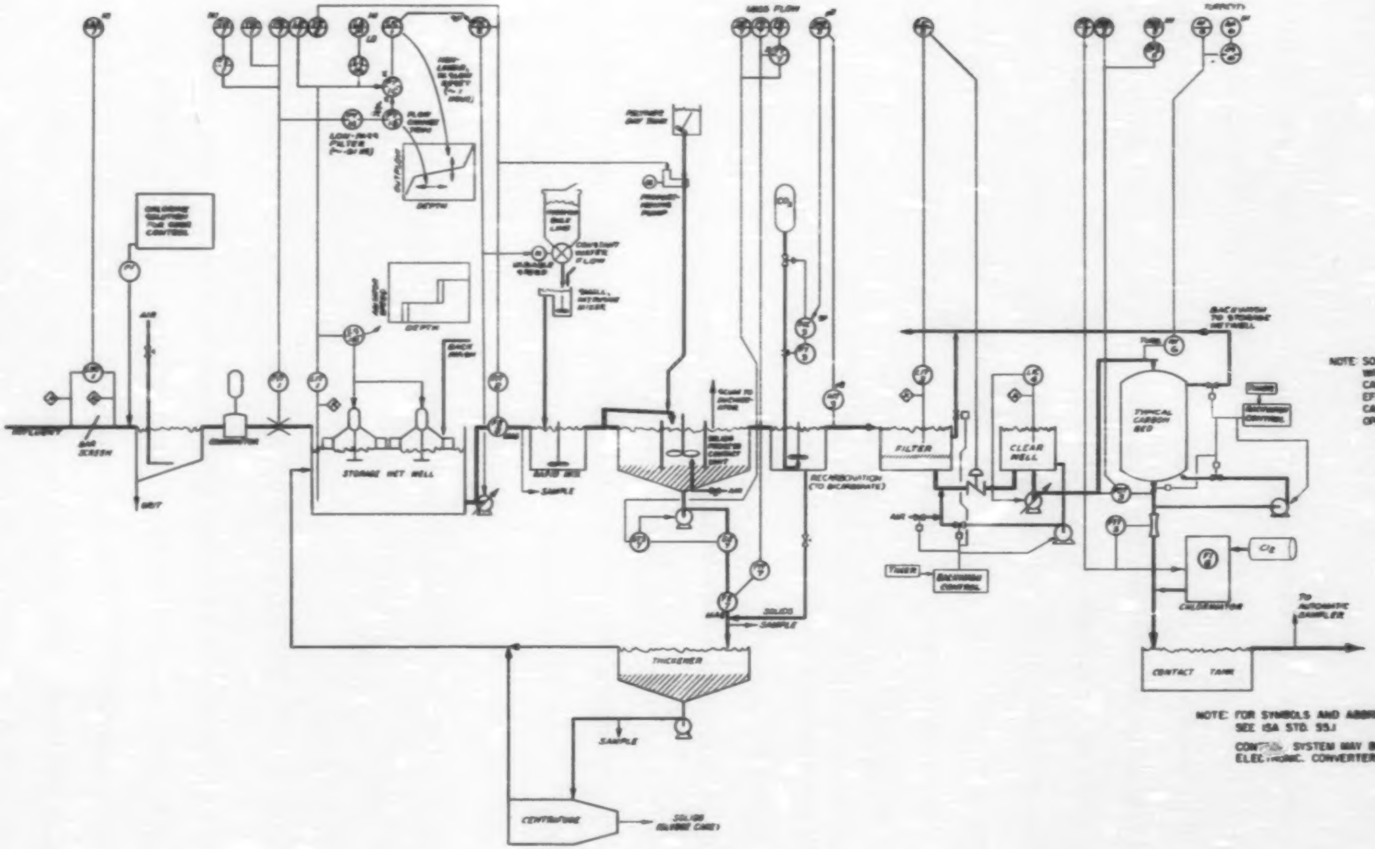
ONE MGD PHYSICAL/CHEMICAL PLANT (Figure 90)

For a discussion of influent monitoring and flow equalization, refer to the previous sections. The physical/chemical plant begins to differ from the activated sludge plant with the feed system for the solids contact unit.

Lime dosage can be feedforward and yet remain accurate and reliable provided delays and nonlinearities are kept small. If the pacing signal for the lime feeder is taken from FIC-2 rather than FE-2, the advantage in time response largely compensates for the unavoidable delay between a signal to the variable-speed lime feeder motor and the moment of lime-slurry delivery into the rapid-mix tank. The lines among the dry lime feeder, the small mixer, and the rapid-mix tank must be short, and the intensive-mix tank must be small. (The time of hydration as well as the time of lime reaction must also be verified as part of the system's design.) The addition of polymers or other coagulants is also controlled by means of a feedforward loop; this has become well-established practice.

Sludge removal is controlled by timers and an in-line density meter (as in the previous example), with the addition of a multiplying relay (FY-7) to indicate the mass flow of solids removal continuously.

A cascaded feedback system for recarbonation is illustrated in Figure 90, where the setpoint of the flow controller (FIC-3) is manipulated by the pH controller (ARC-3).


INFLUENT
BAR SCREEN
GRIT
AIR
COMMINUTOR
STORAGE WET WELL
DEPTH
BACK WASH
MGD FLOW
SAMPLE
CO_2
RECARBONATION (TO BICARBONATE)
FILTER
CLEAR WELL
TIMER
BACKWASH CONTROL
BACKWASH TO STORAGE WETWELL
TYPICAL CARBON BED
TURBIDITY
Cl_2
CHLORINATOR
TO AUTOMATIC SAMPLER
CONTACT TANK
SOLIDS
THICKENER
CENTRIFUGE
SOLIDS (SLUDGE CAKE)

NOTE: SOURCE OF BACKWASH WATER FOR ANY GIVEN CARBON FILTER IS EFFLUENT FROM OTHER CARBON FILTERS IN OPERATION.

NOTE: FOR SYMBOLS AND ABBREVIATIONS, SEE ISA STD. S5.1

CONTROL SYSTEM MAY BE PNEUMATIC OR ELECTRONIC. CONVERTERS ARE NOT SHOWN.

Figure 90. One mgd physical/chemical wastewater-treatment plant, instrumentation diagram

A submerged combustion unit could be used instead but, while this might eliminate the need for a mechanical mixer, it could also cause a possible decrease in response and flexibility.

The filtration system is conventional. Although only one is shown, several filters are employed, each with its own control valve, and all feed a common clear well. Simple feedback controls pace the filtration rate and pump speed to accommodate the input, while the filter condition is indicated for each filter by LDI-3. The backwash is initiated either manually or by a timer (in a system not shown in detail in Figure 90. Overall filtration efficiency is shown by a turbidity analyzer (AI-6), with alarm AA-6 to warn of extreme values. The AT-6 output could also be recorded to indicate trends continuously. (The need for AT-6 results from the requirements of the carbon beds for a feed free of solids.)

Only one of several carbon beds is shown. These beds are connected either in series or parallel. Gross plugging and plugging trends are recorded and alarmed by PDR-5 and PDA-5. If a suitable automatic water quality analyzer were commercially available, this would be the place to use it but, with the present state of the art, carbon bed control and backwashing can be controlled manually, or automatically by a programmed timer, based on the laboratory tests of samples collected manually in whole or in part.

The effluent chlorination and control of the thickener, centrifuge, etc. are not shown in Figure 90.

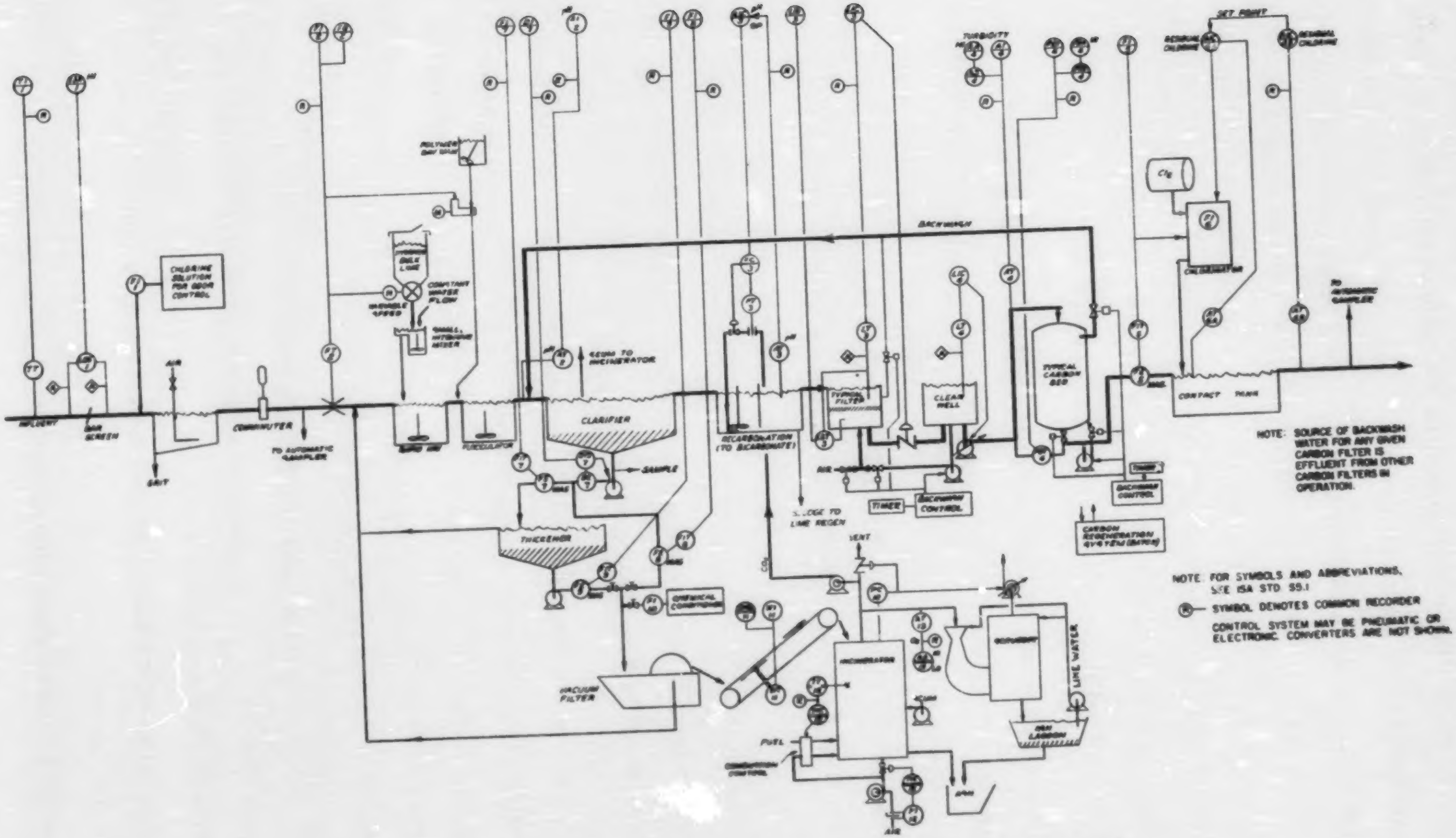
TEN MGD PHYSICAL/CHEMICAL PLANT (Figure 91)

The instrumentation for this complicated and extensive facility is designed to provide the most effective, useful, and reliable performance, as well as the production of meaningful data. Centralized control (in one or several centers) of some 10 to 12 unit operations is assumed. It is further assumed that: 1) no process intelligence at frequencies greater than 0.1 Hz or overall accuracy greater than 2% need be collected, and 2) in many cases, the interaction of measurable values, as well as the same variables measured singly, represents desirable data. Consequently, considering the cost per point of obtaining useful information, multi-channel (sampling type) self-balancing potentiometric records are least costly and thus are recommended for recording up to six points per recorder. One alternative is to use a computerized data collection system, but this is slightly more expensive (in both capital and operating cost) and is also somewhat less reliable overall, so that its selection must hinge on other factors such as increased flexibility and automatic data processing.

The operation of the water purification processes is straightforward and has been described above. Two residual chlorine analyzers are shown, arranged in cascaded feedback loops to ensure an adequate final residual at all times. The savings in chlorine provided by the second analyzer (AT-6B) plus the assurance of meeting effluent standards justifies the cost. These analyzers have been well proven in wastewater service. Although backwashings, thickener overflows, and vacuum filtrates are returned to the head of the plant, there is no sludge recirculation. Control of the thickener and vacuum filter is assumed to be conventional as supplied by the equipment manufacturer, and it is described elsewhere.

The output of the vacuum filter battery is indicated, recorded, and totalized conveniently and reliably from the conveyor belt scale (WT-11). This information provides a good performance index for management, as well as a convenient and responsive indication of filter performance for the operator (he is no longer required to watch every filter so often, once he has put it into service).

The operation of the incinerator and scrubber is merely indicated since these systems are usually provided as complete packaged systems. However, reliable flue gas oxygen analyzers with sampling systems are now commercially available. These operate under plant conditions with little or no maintenance and, in addition, provide an output so dependable that an operator can rely on the information to set the fuel/air ratio to his burners so as to ensure their utmost performance and efficiency.


INFLUENT
BAR SCREEN
GRIT
AIR
COMMINUTER
TO AUTOMATIC SAMPLER
CHLORINE SOLUTION FOR ODOR CONTROL
POLYMER DAY TANK
CONSTANT WATER FLOW
VARIABLE SPEED
SMALL INTENSIVE MIXER
FLOCCULATION
CLARIFIER
SCUM TO INCINERATOR
SAMPLE
THICKENER
CHEMICAL CONDITIONING
VACUUM FILTER
RECARBONATION (TO BICARBONATE)
SLUDGE TO LIME REGEN
VENT
TYPICAL FILTER
CLEAR WELL
TIMER
BACKWASH CONTROL
BACKWASH
TURBIDITY
TYPICAL CARBON BED
CARBON REGENERATION SYSTEM (BATCH)
Cl2
CHLORINATOR
RESIDUAL CHLORINE
SET POINT
CONTACT TANK
TO AUTOMATIC SAMPLER
INCINERATOR
SCRUBBER
LIME WATER
ASH LAGOON
FUEL
COMBUSTION CONTROL
ASH

NOTE: SOURCE OF BACKWASH WATER FOR ANY GIVEN CARBON FILTER IS EFFLUENT FROM OTHER CARBON FILTERS IN OPERATION.

NOTE: FOR SYMBOLS AND ABBREVIATIONS, SEE ISA STD S5.1

Ⓡ— SYMBOL DENOTES COMMON RECORDER

CONTROL SYSTEM MAY BE PNEUMATIC OR ELECTRONIC. CONVERTERS ARE NOT SHOWN.

Figure 91. Ten mgd physical/chemical wastewater-treatment plant, instrumentation diagram

APPENDIX A

FOOD-TO-MICROORGANISM CONTROL DETAIL

SOLIDS BALANCE AROUND AERATION TANK

$$Q_R C_R + YQF - K_d VC_A - (Q + Q_R)C_A = V \frac{dC_A}{dt}$$

$$\frac{dC_A}{dt} = \alpha - \beta C_A$$

where

$$\alpha = \frac{Q_R C_R + YQF}{V}$$

$$\beta = \frac{Q + Q_R + K_d V}{V} \qquad (1)$$

Solution is

$$C_A = \frac{\alpha}{\beta} - \left(\frac{\alpha}{\beta} - C_{A0} \right) e^{-\beta t}$$

$$= \theta - (\theta - C_{A0})e^{-\beta t} \qquad (2)$$

where

$$\theta = \frac{Q_R C_R + YQF}{Q + Q_R + k_d V} \qquad (3)$$

C_{A0} = initial value

To get aeration tank into the desired F/M ratio at time t, desired

$$C_A = C_{AD} = \frac{GFQ}{V} \tag{4}$$

CONTROL PROGRAM FOR Q_R

Say t = 1 hour = 1/24 day

1. On the hour read

$$\underbrace{Q,\ Q_R,\ Q_S}_{\text{MGD}} \qquad \underbrace{F,\ C_R\ C_{AD}\ C_S}_{\text{lb/MG}}$$

2. Assume $Q_R = 0.0$ and compute

C_{AD} equation (4)

β equation (1)

θ equation (3)

C_A equation (2)

If computed $C_A > C_{AD}$, make $Q_R = 0$

If $C_A < C_{AD}$ compute required Q_R as follows:

1. Assume trial value of Q_R and compute β from equation (1)
2. Compute required θ from

$$\theta = \frac{C_{AD} - C_{A0}\, e^{-\beta t}}{1 - e^{-\beta t}} \tag{5}$$

3. Compute required Q_R from

$$Q_R = \frac{\theta(Q + K_d V) - YQF}{C_R - \theta} \tag{6}$$

4. With new value of Q_R repeat the computation until Q_R converges.

5. Change Q_R to computed value.

PROGRAM FOR Q_S

Rate of flow of solids from aeration tank

$$= (Q + Q_R)\,\bar{C}_A$$

$\bar{C}_A$ = average value for interval

$$= \frac{C_{A0} + C_A}{2}$$

To remove solids from final tank at same rate, make

$$Q_S = \frac{(Q + Q_R)\,\bar{C}_A}{C_S} \tag{7}$$

The value of C_S will not be independent of Q_S but will depend upon Q_S and the settling qualities of the sludge. The problem is to maximize C_S and thereby minimize Q_S without making $Q_S\,C_S < (Q + Q_R)\,\bar{C}_A$.

If $Q_S\,C_S < (Q + Q_R)\,\bar{C}_A$, sludge will be accumulating in the final tank and Q_S should be increased.

If $Q_S\,C_S > (Q + Q_R)\,C_A$, excess water is being pumped from final tank into sludge storage, and Q_S should be decreased.

VALUE OF SLUDGE STORAGE

Increase in sludge storage volume during interval = $\Delta V_{SS} = (Q_S - Q_R)\Delta t$

APPENDIX B

NOTATION FOR INSTRUMENT CONTROL LOOPS

The meaning of the instrument symbols can be determined using the information contained in both Figure B-1 and Table B-1. The abbreviations used in the symbols consist of a combination of two or more letters. The first letter indicates the measured variable. The second and third letters are modifiers and instrument function indicaters. The numbers that may be found below the identifying letters are the loop numbers as assigned by the engineer. Detailed information concerning this coding system is contained in Figure B-1 and Table B-1.

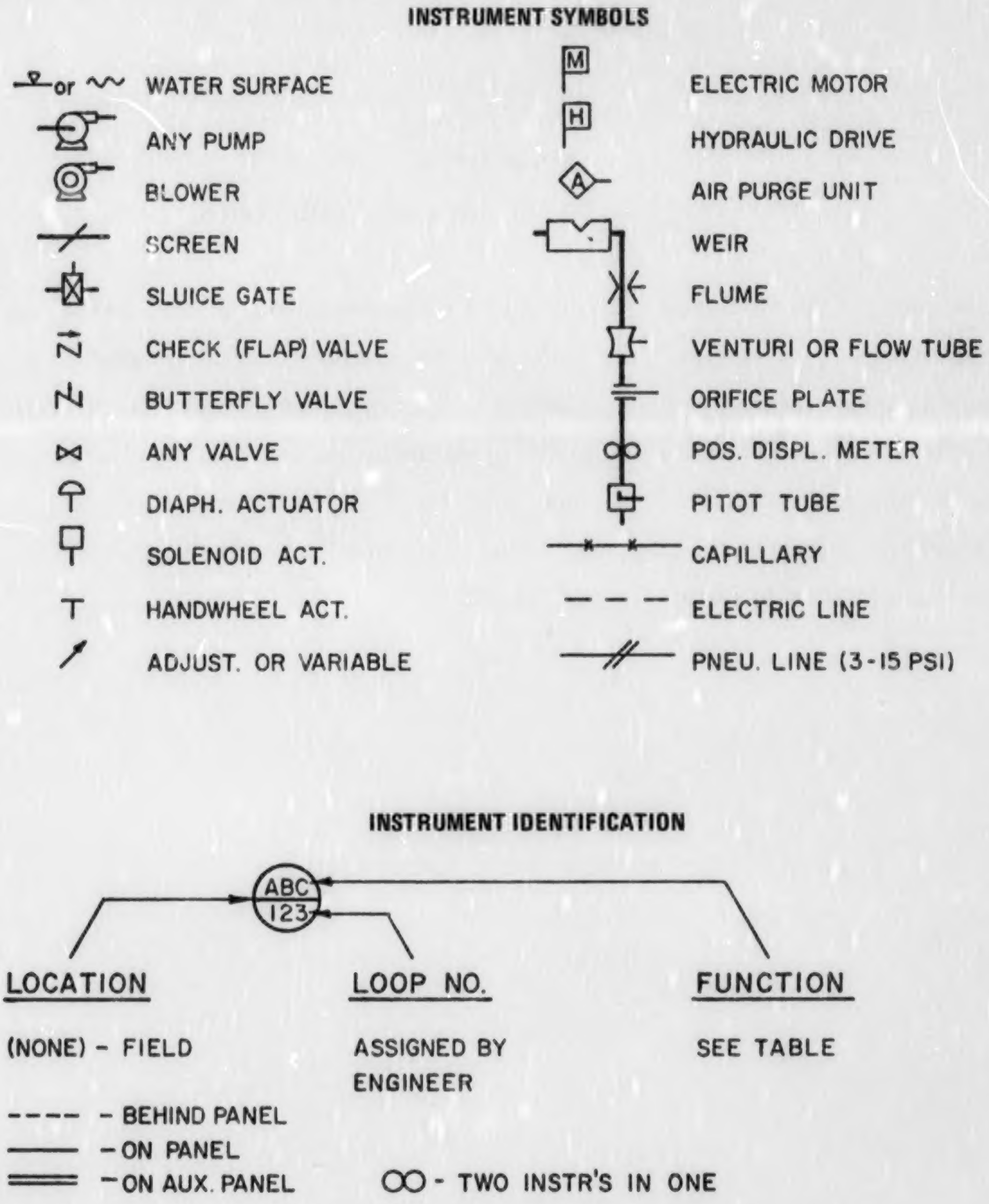

INSTRUMENT SYMBOLS
or
WATER SURFACE
ANY PUMP
BLOWER
SCREEN
SLUICE GATE
CHECK (FLAP) VALVE
BUTTERFLY VALVE
ANY VALVE
DIAPH. ACTUATOR
SOLENOID ACT.
HANDWHEEL ACT.
ADJUST. OR VARIABLE
M
ELECTRIC MOTOR
H
HYDRAULIC DRIVE
A
AIR PURGE UNIT
WEIR
FLUME
VENTURI OR FLOW TUBE
ORIFICE PLATE
POS. DISPL. METER
PITOT TUBE
CAPILLARY
ELECTRIC LINE
PNEU. LINE (3-15 PSI)
INSTRUMENT IDENTIFICATION
ABC
123
LOCATION
LOOP NO.
FUNCTION
(NONE) - FIELD
ASSIGNED BY ENGINEER
SEE TABLE
- BEHIND PANEL
- ON PANEL
- ON AUX. PANEL
- TWO INSTR'S IN ONE

Figure B-1. ISA symbols

Table B-1. INSTRUMENT ABBREVIATIONS

IDENTIFICATION LETTERS
Typical tag for Indicating Flow Controller FIC-3A;

F	I	C	3	A
first letter	second letter	last letter	loop number	suffix

	Measured or Initiated Variable	Modifier	Instrument Function
A	Analysis (1)	Alarm	Alarm
B	Burner (flame)	Special (3)	Special (3)
C	Conductivity	—	Controller
D	Density or Sp. Gr.	Differential	—
E	Electric (General)	Element (measuring)	Element
F	Flow	Ratio (2)	—
H	Hand (Manually Initiated)	—	High
I	—	Indicator	Indicator
J	Power	—	—
K	Time or Time Schedule	—	Control Station
L	Level	Light (Pilot)	Low or Light
M	Moisture or Humidity	—	—
N	Special (3)	Special (3)	Special (3)
P	Pressure (Vac.)	—	—
Q	—	Totalizer	Totalizer
R	Running (Status)	Recorder (4)	Recorder (4)
S	Speed	Switch	Switch
T	Temperature	—	Transmitter
U	Multi-use	—	—
V	—	—	Valve
W	Weight, Force or Torque	Well	Well
X	Special (3)	Special (3)	Special (3)
Y	—	—	Relay or Computer
Z	Position	Special (3)	—

(1) Type of analysis to be defined outside balloon as: pH, ORP, D. O. (dissolved oxygen), R. C. (residual chlorine), TURB (turbidity), etc.

(2) As a modifying letter to designate (fraction) ratio, i. e., FFIC-Flow Ratio Indicating Controller.

(3) As defined in Instrument List of each job.

(4) Or printer.

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